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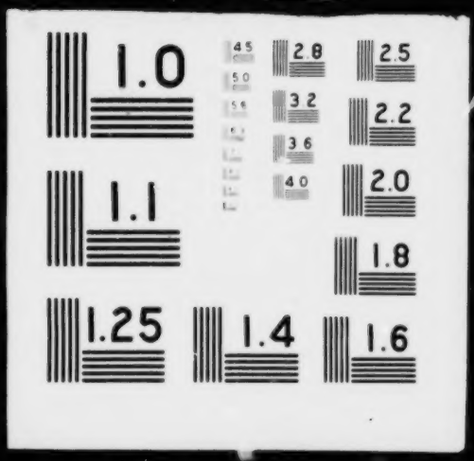

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ENVIRONMENTAL HEALTH SERIES
Water Supply
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ULTIMATE DISPOSAL
of Advanced-Treatment Waste
Part 1. WET OXIDATION
Part 2. INCINERATION

AWTR-3

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U.S. DEPARTMENT OF HEALTH,
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Public Health Service

ULTIMATE DISPOSAL BY WET OXIDATION AND INCINERATION

ULTIMATE DISPOSAL OF ADVANCED-TREATMENT WASTE

Part 1. Wet Oxidation

Part 2. Incineration

by

Louis Koenig

for

The Advanced Waste Treatment Research Program
Basic and Applied Sciences Branch
Robert A. Taft Sanitary Engineering Center

A combination of reports submitted in fulfillment of contracts between the Public Health Service and Louis Koenig-Research, San Antonio, Texas, (SAph 78083 and Purchase Order No. 680) Parts 1 and 2 were submitted in July 1961 and February 1962, respectively.

U. S. DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
Public Health Service
Division of Water Supply and Pollution Control

Cincinnati, Ohio
October 1963

Advanced Waste Treatment Research Publications

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AWTR-1 Summary Report, June 1960 - December 1961 W62-9

AWTR-2 Preliminary Appraisal of Advanced Waste Treatment Processes W62-24

AWTR-3 Ultimate Disposal of Advanced Treatment Waste 999-WP-3

Part 1. Wet Oxidation
Part 2. Incineration

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Public Health Service Publication No. 999-WP-3

ADVANCED WASTE TREATMENT RESEARCH PROGRAM

The Advanced Waste Treatment Research Program of the Public Health Service has two ultimate goals: One is to help abate our Nation's growing water pollution problems and the other, more startling in concept, is to renovate waste water for direct and deliberate reuse.

Conventional water and waste treatment practices have little or no effect on many simple inorganic salts and permit an ever-increasing number and amount of highly complex synthetic organic wastes to contaminate drinking water supplies. If we are to protect these water supplies for the future, it is mandatory that new technologies for water and waste treatment be developed.

Advanced Waste Treatment may be looked upon as a two-step process: (1) separating concentrated contaminants from the purified water "product" and (2) disposing of the concentrated contaminants in a way that will render them forever innocuous. This "permanent disposal" of separated wastes is essential so that useable surface or ground waters will not be contaminated and so that the same contaminants need not be removed again and again from water supplies.

In one step, then, advanced waste treatment, i.e., water renovation, could alleviate both water pollution and water supply problems - problems of increasing concern both in this country and in the rest of the world. The importance of advanced waste treatment was recognized by Congress in its 1961 amendments to the Federal Water Pollution Control Act (Public Law 87-88). This act directs the Secretary of Health, Education, and Welfare "... to develop and demonstrate practicable means of treating municipal sewage and other water-borne wastes to remove the maximum possible amounts of physical, chemical, and biological pollutants in order to restore and maintain the maximum amount of the Nation's water at a quality suitable for reuse."

It is too early to predict accurately the cost of advanced waste treatment technology, but the need is inevitable. Because of population increase and industrial expansion, we are approaching the time when our rivers and streams will no longer be able to assimilate our wastes and when there will be no more developable supplies of fresh water. Whatever the cost of this technology, one thing is certain--it must be the very lowest science and engineering can achieve. Anything more will impose an inexcusable, perhaps intolerable burden on our country's economic growth and on its ability to compete in tomorrow's world.

The author of this report, Dr. Louis Koenig, has specialized in waste treatment and disposal and their economics for several years. He has been involved in research projects on the injection of natural brines in the Arkansas - Red River Basin and on several processes for disposing of saline water conversion wastes. He has also studied, for the AWTR Program, injection into underground formations, soil spreading, and placement in underground cavities as ultimate disposal techniques. Reports on the latter studies will be published later.

PREFACE

Under the Advanced Waste Treatment Research Program, the U.S. Public Health Service is engaged in a study of processes which might be used to treat various waters or waste waters in order to make them reusable. Some of the processes would completely remove or destroy the contaminants, whereas others would leave the original contaminants as a waste to be disposed of. The processes for dealing with this waste are termed "Ultimate Disposal." Ultimate disposal means the treatment, handling, or placement of a waste in such a manner that it never again comes in contact with human activity other than, in certain instances, a very small surveillance or containment activity. These reports present the results of preliminary studies of two methods of ultimate disposal of the organic contaminants, namely wet combustion and incineration. The work was instituted under contract No. SAph 78083 and Purchase Order No. 1680 between the Robert A. Taft Sanitary Engineering Center and Louis Koenig-Research, covering a scope as follows:

> "Assess the physical, natural and economic parameters which might bear on the costs, physical magnitudes, and general feasibility of ultimate disposal of advanced treatment wastes by wet combustion and incineration. Determine by preliminary cost estimation for reasonable ranges of these parameters, the costs and physical magnitudes involved especially as these are influenced by volume and concentration of waste. Point out areas in which further research or data collection would be necessary in order to make a similar estimate of the next order of accuracy."

This is part of a planned series of studies which will cover a number of other possible methods of ultimate disposal and their interrelations with the advanced treatment processes themselves.

Responsibility for the overall report is that of the author. Work on plant design and cost estimation was performed by Graham B. Knight, Consulting Chemical Engineer, of San Antonio. Special acknowledgement is due to several individuals who kindly supplied data and information relative to costs and design in the incineration study. Among these are A. W. Grahame, Jr., Nichols Engineering & Research Corp., and Dr. W. H. Gauvin, Alfred Assaly, Jean J. O. Gravel, and C. F. B. Stevens, Pulp and Paper Research Institute of Canada.

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ABSTRACT

Cost estimates are made for ultimate disposal of the concentrated contaminants resulting from complete renovation of a municipal waste water. The processes investigated are wet oxidation (Zimmerman Process) and incineration.

For wet oxidation, plant costs vary from $4 to $80 per gallon of concentrate per day capacity (10,000,000 to 1,000 gallons per day). Operating costs vary from $1.40 to $35 per thousand gallons of feed. Pre-evaporation could reduce costs. Fixed costs constitute a large proportion of total costs. A theory of the reaction mechanism is proposed (8 figures, 19 references).

For incineration, estimates of the cost of pre-evaporation to 30 percent solids and hauling of the ash were included. A flash dryer-Nichols Herreshoff furnace combination was selected. Costs per 1,000 gallons of concentrate varied from $0.57 (10,000,000 gallons per day, 300 ppm) to $68 (1,000 gallons per day, 300,000 ppm). Corresponding unit investment costs were $1.53 and $108 per gallon per day capacity (11 figures, 30 references).

ULTIMATE DISPOSAL

BY

WET OXIDATION

Part 1.

WET OXIDATION

Summary

In some of the processes being investigated under the Advanced Waste Treatment Research Program of the U. S. Public Health Service, water is extracted from the raw materials or contaminants are removed therefrom to yield the ultimate product and a waste stream consisting of the contaminants and possibly some of the reagents in a more or less concentrated form. A preliminary process design and cost estimate was made for ultimate disposal of the organic contaminant in such wastes by wet oxidation.

Nine hypothetical cases covered the flow ranges from 1,000 to 10 million gpd and organic contaminant concentration ranges from 100 to 100,000 ppm. Reactor design data (for 80% removal of COD), process cost data, and other process parameters were developed from pilot plant data reported in the literature, from design data for plants not yet in operation, and from labatory data and other information obtained from the developers of the wet oxidation process. The least secure portion of reactor design data involves the extrapolation of performance from initial COD concentrations of around 13,000 ppm to the lowest concentration used in the study, 200 ppm. The cost of settling, filtering, and hauling the suspended solids in the effluent was not included because no data were available on the amounts to be expected from advanced treatment wastes.

A theory of the reaction mechanism, based on available data, is proposed to explain some of the peculiarities of the reaction. The postulates include: that the overall reaction consists of a sequence in which intermediate products of lower molecular weight are successively formed; that the reaction rate of each intermediate increases with increasing molecular chain length of the intermediate; and that the size of the molecular fragments produced in each step decreases as the temperature increases. These postulates lead to qualitative predictions concerning the amount of acids in the effluent liquor, as confirmed by experimental data.

The process design shows that for a fixed ratio of inorganics to organics in the feed, here 2.0, there is an upper limit of organic concentration which can be wet oxidized without over-

saturation with respect to the inorganics. This concentration is 60,000 ppm organics (i.e. 6%) for the conditions chosen.

Plant costs per gallon per day of capability vary from about $4 at 10 mgd to $80 at 1,000 gpd, the effect of feed concentration being minor. All costs are adjusted to a Marshall & Stevens Chemical Process Industries Equipment Cost Index of 240. Disposal costs per thousand gallons vary from about $1.40 at 10 mgd to $35 at 1,000 gpd, initial concentration being of minor influence. Disposal cost in dollars per day is practically independent of the tons per day of organics in the feed at any given gpd feed rate. At the upper limiting concentration the cost per ton of organics oxidized varies from $141 at the lowest flow rate, 1,000 gpd, to about $6 at 10 million gpd. Charts are presented by which various interpolations of the above unit cost figures may be made. The fixed costs constitute a very large proportion of the total disposal costs; thus, for an established plant the daily disposal cost is rather insensitive to the actual throughput rate.

As an example for practical use there is computed the costs for wet oxidation processes for a 60,000-population community producing 10 mgd. Conventional secondary treatment would cost about $700 per day. Wet oxidation of the sludge from this plant, replacing the present sludge disposal method, would cost $200 per day. Direct wet oxidation of the raw sewage or of the secondary effluent would cost $13,000 per day.

As an example for the guidance of advanced treatment research there is computed the daily cost of wet oxidation for advanced treatment processes under the above conditions for various percent recoveries of water. That very high recoveries in the advanced treatment process are necessary in order to keep the ultimate disposal costs within reasonable bounds is shown in the data that at 99.83 percent recovery ultimate disposal would cost $15/million gallons of water produced, while at 96.67 percent recovery the cost would be $91.

Among areas suggested for further investigation are the kinetics and mechanism of the reaction, the possibility of using oxidation catalysts, and methods of disposal of the organics still residual from wet oxidation. The latter methods include biological oxidation and chemical precipitation.

Derivation of Reactor Design Data

AVAILABLE DATA

A minimum of reactor design data were available for the preparation of this analysis. It comprised data from two sets of

pilot plant experiments[6] for the West-Southwest pilot plant and for the Rothschild pilot plant; three sets of partial design data for commercial plants, the Chicago, the Wheeling, and the Wausau[3]; and two sets of curves showing data which was said to have been obtained from rocking autoclave experiments. (Table 1)[1, 2, 5, 6]

Table 1. WET OXIDATION REACTOR DESIGN DATA

Plant	Dry solids, tons/day	Dry solids, %	COD, g/l	Feed gpm	Labor per shift, men	Temperature, F°	Pressure, psi	Cost for power	Capital cost, $million	COD Fraction removed
Chicago	4 x 50	3	40	1200	4		1800	$1.00/ton	12	.70
Wheeling	5.6	6	80		1/2			0	0.45	
Wausau	3.5	5			1/2			0	0.325	
Borregaard	2 x 150		140	700		590	2250	yields 240,000 lb/hr steam	5.00[a]	.90
5000 population equivalent	0.5				2 hr/day[b]		500	not recovered		
estimated	1.0				2 hr/day[b]			not recovered	0.10	

[a] This figure, which is for Norway, would be much higher in the United States.

[b] These plants operate 20 hours per day per batch. The sewage plant operator devotes about 2 hrs per day in a 4-hour period refilling.

The essential parameters among which relations are sought are initial COD concentration in feed liquor, x_0; fraction of COD concentration removed, f; reactor temperature, T_R; and reactor detention time, t_R. These together with other symbols and definitions used in this report are shown below. The ranges in these essential parameters over which data were available were:

Range of all data		Range of pilot plant data	
t_R	0 to 200 minutes	t_R	26 to 200 minutes
T_R	25° to 300°C	T_R	488° to 541°F
x_0	13 to 98 mg/l	x_0	43.3 to 72.2 mg/l
		f	0.723 to 0.864

LIST OF SYMBOLS USED

f - fractional change in COD concentration $= \frac{x_0 - x_f}{x_0}$

$= (\sim f_{corr}$ when no H_2O is evaporated)

$f_{60\ gpl}$ - fractional change starting from feed of $x_0 = 60$

$f_{1\ hr}$ - fractional change when t_R = 60 minutes

f_{∞} - fractional change after an extended time

f_{corr} - fraction of entering weight of COD removed, (1.06 to 1.09 x f in range of experimental conditions)

k - rate constant, $min^{-1}\ (g/l^{(1-n)}$

n - exponent, order of reaction

q - liquid feed rate, gallons per minute at ambient conditions

t - time, minutes

t_f - final time, minutes

t_o - initial time, minutes

t_R - reactor time (reciprocal space velocity), minutes - $V_R/q = t_f - t_o$,

T_R - temperature of reaction medium, °C if not specified °F

V_R - volume of reactor, gallons

x - concentration of COD, grams per liter

x_0 - initial concentration of COD in feed liquor ambient, g/l

x_f - final concentration of COD in effluent liquor ambient, g/l

ATTEMPT TO DEVELOP A KINETIC EQUATION

The problem was to develop a relationship among the above parameters in the range of the available data, i.e., a relationship that may be used to predict performance outside this range of data, particularly at much lower initial concentration, x_0, corresponding to the three design cases:

	Organic, ppm	Corresponding COD, g/l
High	100,000	200
Intermediate	3,000	6
Low	100	0.2

Some data of uncertain origin and application were available relating x_0 and t_R for T_R = 271°C and f = 0.80, together with a similar estimate of a single point for f = 0.70. The data for both f = 0.80 and f = 0.70 were fitted well by a simple kinetic equation:

$$\frac{dx}{dt} = kx^n = -0.000664\ x^{2.8462} \qquad (1)$$

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in the integrated form:

$$x_f^{-1.8462} - x_0^{-1.8462} = 0.001225\, t_R \qquad (2)$$

However, when the t_R's were developed for the pilot plant data and the plant design data, they were found to be many times higher than the t_R's corresponding to this equation.

The simple kinetic equation, Equation (1), requires n=6 or more to reproduce the pilot plant and rocking autoclave data of f vs t_R down to the lowest t_R available, about t_R =10. The deviation is that the equation falls off in f too much at low t_R. These data apply for x_0=45, the conditions of the pilot plant experiments. However, even when the best fit of n=5 is used for x_0 =45, then for low x_0's, the equation is much too low at all t_R's. In other words the simple kinetic equation cannot simultaneously represent the pilot plant and rocking autoclave data for both reactor time and initial concentration.

The possibility should be metioned that the reactor times corresponding to the above Equation (1) may have been reactor times computed for the total volume of liquor, noncondensable gas and steam at reactor conditions corresponding to the rate of liquor feed at ambient conditions. The volume of steam, noncondensible gas, and liquor is many times the volume of the ambient liquor feed. For example, under the conditions of Run No. 5 at the West-Southwest pilot plant, 488°F, 1220 psi, and 1.59 pounds of air per gallon of liquor feed, the volume of reactor fluid per gallon of ambient feed liquor would be 8.76. Even higher values are found for some of the other experiments. Possibly these steam-gas-liquor volumes were used in computing reactor times for the data mentioned above. But the kinetic data suggest that the important parameter is liquor retention time. The density of the gas in the reactor even at these high pressures is still only about 1/15th that of the liquor so that the rate of bubble rise is very high compared to the rate of liquor rise. The reactor must be considered as being filled with liquor except for the head space, with the bulk of the liquor containing only a small amount of gas at any one instant. Thus, liquor retention time must be used for reactor time if one is to correlate between rocking autoclave experiments and continuous flow pilot plant experiments.

One further correction along this line might have been made. The specific volume of the liquid at reactor conditions is about 0.0201 cubic foot per pound compared to a specific volume at ambient conditions of 0.016. Thus the actual volume at reactor

conditions is 25 percent greater than at ambeint conditions and the reactor times should strictly be correspondingly decreased for the continuous pilot plant data. Compensating however is the fact that the autoclave times at reactor temperature are also lower than indicated, allowing for the time required to reach reactor temperature.

EMPIRICAL DEVELOPMENT OF RELATIONS

In the previous section it is concluded that the data are not amenable to simple kinetic treatment and the relations must thus be developed empirically.

In Figure 1 the relations between fraction removed and reactor time are shown. The solid lines are taken from the rocking autoclave experiments. The points are the plant design times and that portion of the Rothschild and West-Southwest pilot plants data at 271 ± 2°C, taken as a constant temperature. This temperature was chosen because it corresponded to the data for the t_R-x_0 relationship described above. It is seen that the 271° pilot plant data corresponds approximately with the 250° rocking autoclave data, and that at this temperature and above, there is very little variation of f with t_R. The data are shown in Table 2.

The other available rocking autoclave data [5,6] show curves from which a relation between f and x_0 at t_R = 60 minutes can be calculated. Figure 2 has been developed from a reading of these curves using as x_0 the COD value extrapolated to 0° from the curves. The data for the Imhoff sludge are divergent from the rest of the data. It is seen that down to x_0 = 13, $f_{1\ hr}$ changes very little with x_0 at 250° and 300°. The one point shown for the Rothschild pilot plant at 271° falls within the range of the rocking autoclave data for 250° as previously noted.

From the 250° data in Figures 1 and 2, which are taken to correspond to 271° in the continuous plant, there may be drawn arithmetic corrections for f, which when added to f for 1 hour and a standard x_0 of 60 grams per liter produce the plotted f values at other values of x_0 and of t_R. The corrections to x_0 were taken directly from the cartesian coordinate plot of Figure 2. These corrections were then applied to the pilot plant data to bring it to x_0 = 60 and a second plot of $f_{60\ g/l}$ against log t_R was made for obtaining the adjustments to bring

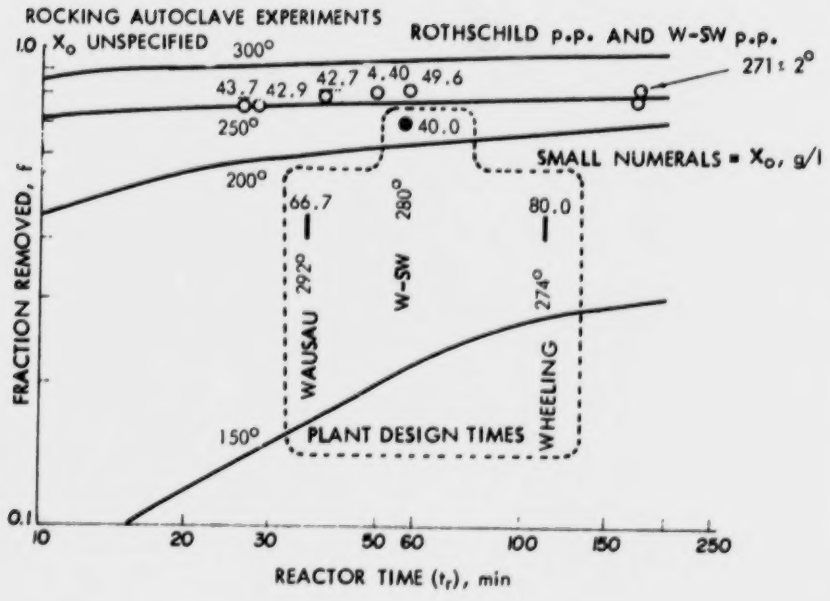

Figure 1. Removal as a function of reactor time at constant temperature.

Table 2. REACTOR TIMES AND OTHER DATA FOR 271° AND f ≈ 0.80

Run number	V_R, gal	q, gpm	t_R, min	T_R, °C	x_0, g/l	f
W-SW pilot plant						
11	705	4.13	171	271	82.9	0.788
17	705	4.08	172	271	60.2	0.815
Rothschild pilot plant						
636-1	98.6	1.67	57.2	273	49.6	0.818
643	98.6	2.00	49.4	271	44.0	0.817
644	98.6	2.58	38.3	273	42.7	0.801
646	98.6	3.52	28.0	272	42.9	0.762
647	96.6	3.80	26.0	271	43.3	0.772
Design values for comparison						
W-SW plant						
(each unit)	15,220	273	56.0	280	40	0.70
Wheeling plant						
	1,680	15.25	110	273	80	[a]
Wausau plant						
	350	9.8	35.8	292	66.7	[a]

[a] Not available.

t_R to the standard 60 minutes. The adjustments were small in f. The West-Southwest pilot plant data for 271° are deviant on this plot, since the excessive reactor times did not yield an increase in f. According to the rocking autoclave data of f vs t_R at the reactor times of the West-Southwest pilot plant, f is still rising slightly, but the West-Southwest pilot plant fraction removed is not correspondingly higher than the fraction removed in the Rothschild pilot plant (at the same reactor temperature). In computation of the adjustment then, the extension of the Rothschild pilot plant data was utilized to draw the curve between the highest t_R of the Rothschild data (at 271°C) and the ultimate fraction removed. The ultimate f was determined by applying to the $f_{1\ hr}$ the ratio $f_\infty / f_{1\ hr}$ obtained from the rocking autoclave experiments at 250°C. The difference being discussed in the above is about 0.05 in f.

From these two adjustment curves, we now have data to correct any x_0 to x_0=60 and any t_R to t_R=1 hour, all applicable at 271°C.

When these two corrections are applied to the pilot plant data at all temperatures, the points appear as in Figure 3. Used to draw the trend line and band width on this figure are the corresponding f_{1hr} at x_0=60 from Figure 2, but transposed upwards in temperature by 21°, bringing the rocking autoclave data into coincidence with the pilot plant data.

Figure 3 adjusted by Figure 2 and Figure 1 was used in this study to produce the fraction removed at any temperature, initial concentration, and reactor time.

The most important extrapolation for our purposes is that of f_{1hr} from an initial COD concentration of 13 grams per liter down to 6 and to 0.2 grams per liter, corresponding to the middle and low concentration cases. It is know that some chemical reactions follow quite different kinetic courses when the dilution becomes extreme. Some inquiry into the reasonableness of this extrapolation is in order. If it is assumed that the average molecular weight of the organic contaminants is 200, then the range of the experimental data from about 13 to about 100 grams per liter COD corresponds to 0.03 to 0.25 molal. The intermediate and low concentrations correspond to 0.05 and 0.0005 molal, the latter is 1/60th the concentration of the lowest experimental concentration. Practically the entire range of the experimental data as well as the desired predictions fall in molalities in which activity coefficients of ions are decreasing with increased concentration following the Debye Hückel theory. At a COD pf 13 each

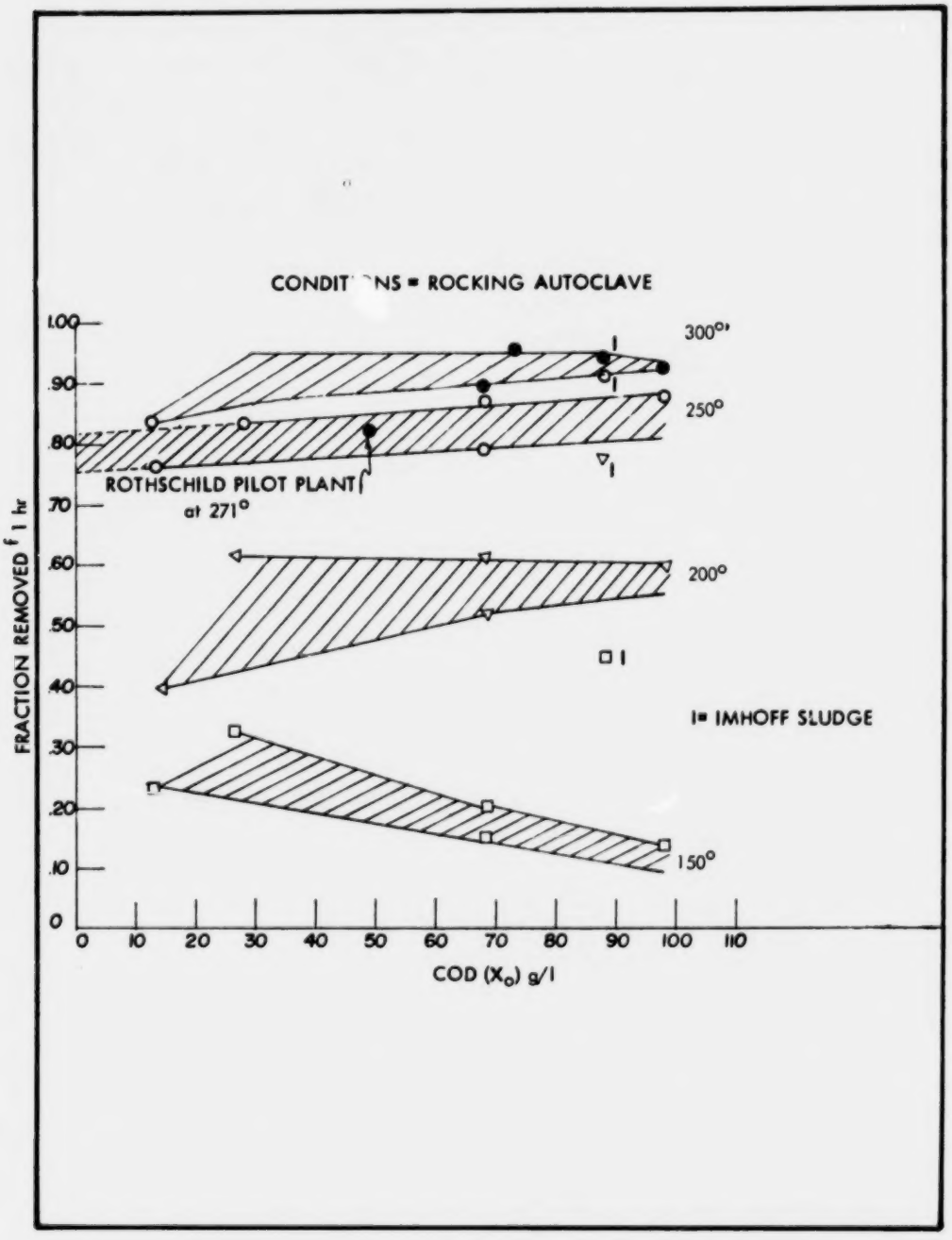

Figure 2. One-hour conversion as function of initial COD at constant temperature.

molecule of 200 molecular weight would be found in 1800 molecules of water, whereas for COD's of 6 and 0.2 the corresponding numbers would be 3600 and 110,000. It would appear that at one molecule in 1800 the molecule is already "lost" so far as the effect of any additional water molecules is concerned. The reaction is stated to be independent of the oxygen pressure and

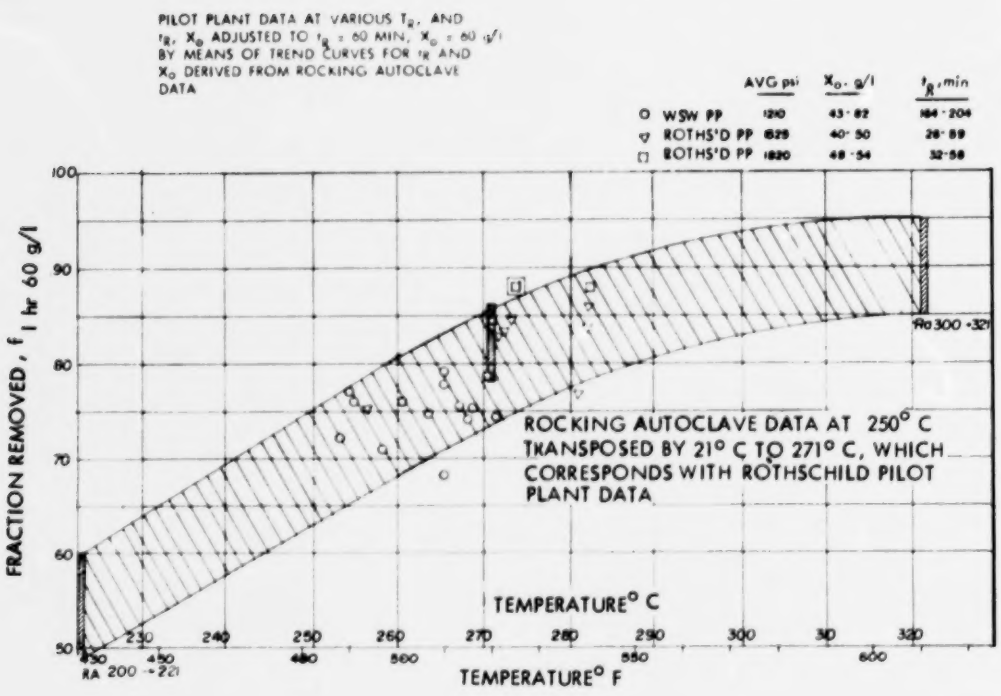
PILOT PLANT DATA AT VARIOUS T_R, AND t_R, X_0 ADJUSTED TO t_R = 60 MIN, X_0 = 60 g/l BY MEANS OF TREND CURVES FOR t_R AND X_0 DERIVED FROM ROCKING AUTOCLAVE DATA
AVG psi
X_0, g/l
t_R, min
WSW PP 1210 43-82 184-204
ROTHS'D PP 1629 40-50 28-89
ROTHS'D PP 1820 48-54 32-58
ROCKING AUTOCLAVE DATA AT 250° C TRANSPOSED BY 21° C TO 271° C, WHICH CORRESPONDS WITH ROTHSCHILD PILOT PLANT DATA
FRACTION REMOVED, f_1 hr 60 g/l
TEMPERATURE° C
TEMPERATURE° F
RA 200 -22
RA 300 -32

Figure 3. Effect of temperature on removal.

this suggests that there is already an excessive amount of oxygen present even at the higher pilot plant concentrations. Accordingly, one could expect that there is also an excess of oxygen at the low concentrations. It is true that according to the simple kinetic equations such as Equation (1) f approaches 0 as x_0 approaches 0.. However, we have already seen that no simple kinetic equation with a reasonable exponent can reproduce the pilot plant and rocking autoclave data over the range of reactor times without representing f as far too small at the lower concentrations. Accordingly, we have already rejected the simple kinetic equation for describing the reaction even in the higher concentrations of the pilot plant and rocking autoclave data. For this reason, the implications of the equation in exactly the deviant region cannot be cited as a bar to extrapolation. In general then, we cannot cite any convincing reasons either for or against the extrapolation from the lowest experimental concentration, 13 grams per liter, to the lowest design concentration, 0.2 grams per liter, and must take the extrapolation therefore as empirically based on the trend of the data from 100 to 13 grams per liter.

Theories of the Mechanism of the Wet Oxidation

The wet oxidation reaction, at least in data available, has some peculiarities which warrant explanation.

The oxidation reaction apparently proceeds to produce lower molecular-weight oxidized compounds typically from initially high-molecular-weight reactants. The end product remaining in the liquor is said to be low-molecular-weight acids, formic, acetic, and such. For these molecules it is known that the carboxyl group strengthens the resistance of the C-C bonds adjacent thereto. Thus, acetic acid is very resistant to oxidation and this indeed is the basis for its manufacture by vapor phase oxidation. It is not unexpected therefore that acetic acid is one of the end products of the wet oxidation. The low-molecular-weight organic acids have some volatility, but the amount of stripping to be anticipated in the concentrations in the product liquor and in the reactor, which can consist at the best of a very imperfect one-stage contactor for distillation, would be inappreciable. Qualitatively, therefore, one would expect that acetic acid as formed would remain in the liquor unoxidized and undistilled. However, acetic acid, which is taken as typical of the low-molecular-weight acids resistant to oxidation, itself has a COD. Therefore, by the analytical methods used, this portion at least of the product of the oxidation is indistinguishable from unreacted reactant. This is the first peculiarity of the data as available.

There are two mechanisms by which COD disappears from the liquor. One of these is the complete oxidation of C and H to CO_2 and H_2O, which have no COD. The other is by partial satisfaction of the COD of intermediate, partially oxidized compounds, which remain in the solution. This means that the residual COD has two components, one the result of some of the reactants being oxidized to the end products and the other the result of the remainder of the reactants being oxidized only partially.

Further characteristics or peculiarities of the reaction are as follows:

1. Oxygen pressure has no effect on the rate or the extent of conversion.

2. Reaction produces CO_2 and H_2O, plus low-molecular-weight acids, e.g. CH_3COOH, HCOOH, etc.
3. Reaction has an initial extremely rapid rate followed somewhat abruptly by a much lower rate.[5,6]
4. Reaction reaches a different ultimate fraction converted at each temperature.
   (Note: This characteristic is not a peculiarity for a reaction reaching an equilibrium point which varies with the temperature. However, in the present case there is no possibility of equilibrium, since the intermediate or final products are very unlikely to recombine to give compounds of higher COD, and this in effect would mean a reduction.)
5. Reaction reaches approximately the same ultimate fraction regardless of initial concentration at high temperatures.
6. Reaction shows a <u>decrease</u> of fraction converted with increase of initial concentration at low temperatures.
7. The fraction converted is a probability function of temperature.
   (Note: This analysis is not described in the report, but was made in the course of the work. The original data[1,2,5,6] on fraction removed as a function of temperature have a form similar to a cumulative probability plot. When plotted as probability function of fraction removed against temperature they approximate a family of straight lines. This is also true for the derived data on ultimate fraction removed.)

The author has made some hypotheses concerning reaction mechanisms which qualitatively serve to explain some of the above peculiarities. These hypotheses are as follows:

1. The over-all reaction consists of a sequential series of intermediate reactions in which lower molecular-weight and possibly partially oxidized intermediate products are formed, and these in

turn are further broken down and oxidized in the next step.

2. The reaction is a chain reaction propagated by some sort of free radical process.
3. The reaction rate, i.e. the lability of the reactants and intermediates, is a function of their molecular chain lengths and increases with increasing molecular chain lengths, or molecular weight.
4. The size of the molecular fragments produced decreases as the temperature, i.e. the severity of attack, increases. (Explanation: At high temperatures the high energy involved fragments the molecule almost entirely into small molecular fragments, typically CO_2, HCOOH, and CH_3COOH. If the original molecule may be considered as a C_{20} molecule, there would be at high temperatures a minimum of say C_8 molecules formed. At low temperatures, however, the C_{20} molecule would break up into, for example, two C_8 molecules, two CO_2, and one CH_3COOH.
5. At each breaking of a molecule at least one CO_2 is formed.

The above hypotheses may be used to explain some of the characteristic features and peculiarities of the reaction.

The fact that the ultimate fraction removed is a function of temperature, increasing with temperature, is explainable as follows. At high temperatures the molecule breaks up almost completely into CO_2 with some residual CH_3COOH, leaving few intermediate size molecules, even for the second step. At low temperatures a small amount of CO_2 and CH_3COOH is formed in the first breakdown but the intermediate size molecules, e.g. C_8, have a lower specific reaction rate by hypothesis No. 3, and at these low temperatures may be completely unreactive. At intermediate temperatures the molecule breaks down into a series of molecules with intermediate amounts of CO_2 and CH_3COOH. These have a lower specific reaction rate than the C_8's of the low temperature case, but the temperature is higher so that the over-all rate of their disappearance is not nil but moderate.

This visualization is verifiable by some of the available experimental facts. If one considers the "small molecules," i.e. CO_2, HCOOH, etc., as a group, the hypotheses demand

that when the production of the low molecule group is large the production of CO_2 and also CH_3COOH is large; whereas, when the production of small molecules is small the production of CO_2 and of CH_3COOH is also small. When the production of CO_2 is large the fraction converted is large, but by the above reasoning, the production of CH_3COOH must also be comparatively large. Thus, when the fraction converted is large, i.e. large production and loss of CO_2, as compared to when it is small, the residual CH_3COOH should make up a large part of the product remaining. When the fraction removed is small, the CH_3COOH (bearing some relationship to the CO_2 produced) should be a small portion of the residual COD, which contains a relatively large amount of intermediate size unoxidized or incompletely oxidized fragments.

The above somewhat surprising deduction is verifiable by the available experimental data, which are shown in Figure 4. The upper portion of the figure shows that, as one might expect intuitively, the amount of volatile acids in the effluent decreases as the fraction removed increases. It must be remembered that these experiments from the Rothschild and the West-Southwest pilot plants have a roughly constant initial concentration of COD. The bottom portion of the figure shows however, the distinct trend of the volatile acids to make up a greater portion of the residual COD, as the fraction removed increases.

The initial rapid rate followed rather abruptly by a much slower rate follows from the hypotheses in this way. The large initial molecules are very labile. The smaller fragments into which they break up have a lower specific reaction rate: thus, as soon as the first step has taken place, the sequential reaction occurs at a much lower rate. That the abruptness of this change increases with increasing temperatures is also explainable by this hypothesis. At the high temperatures the fragments are very small and thus have an extremely low specific reaction rate, compared to the initial molecules. At the lower temperatures, however, the products of the first step are still of moderate size and thus have moderate specific reaction rates. Accordingly, at low temperatures one would expect a slow and smooth transition, whereas at high temperatures one would expect a rapid transition between an extremely high rate and an extremely low rate. This is exactly the nature of the observed curves.

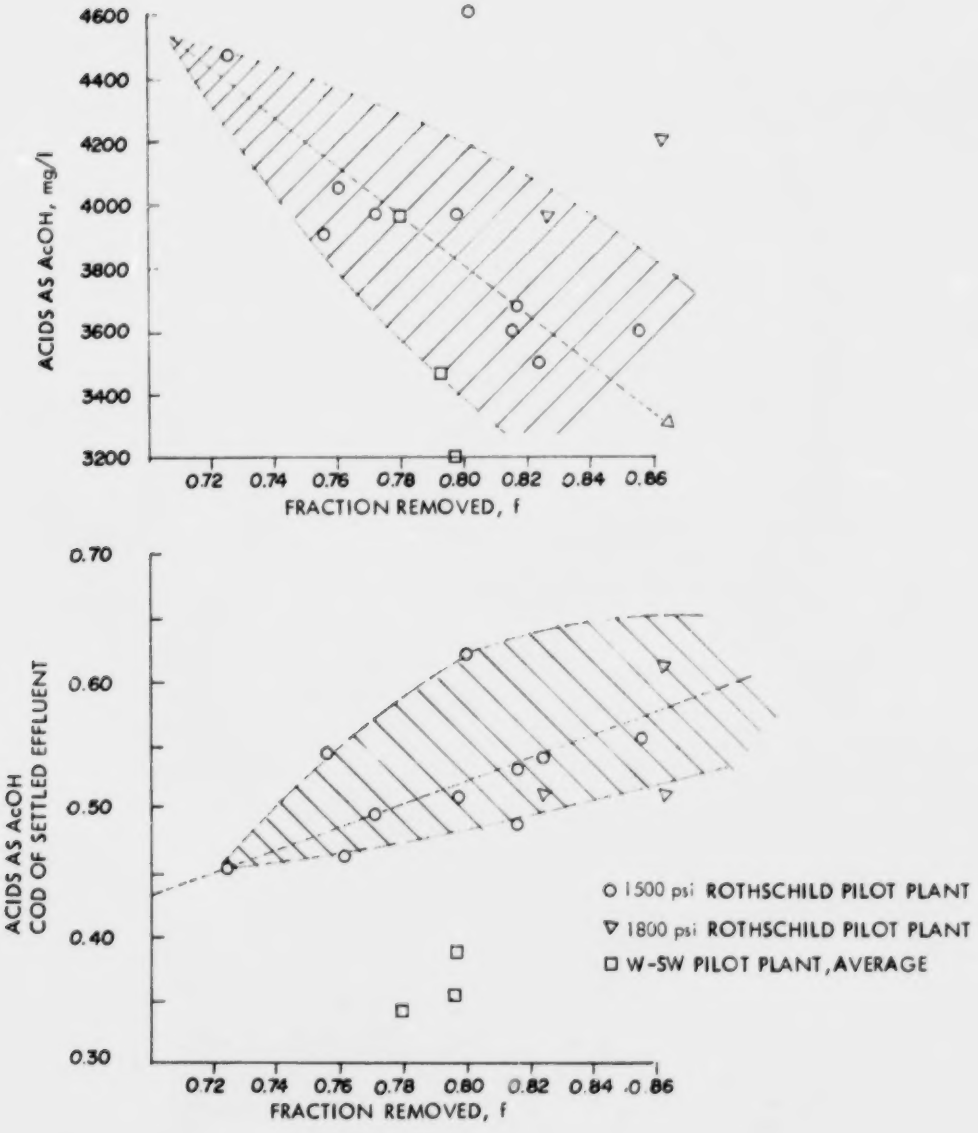

Figure 4. Residual volatile acids as a function of fraction removed.

The fact that the reaction reaches approximately the same ultimate fraction removed regardless of the initial concentration at high temperatures is also explainable by the hypotheses. At high temperatures each molecule is immediately broken up not into intermediate size fragments but into the group of small molecules CO_2, CH_3COOH, etc. The molecule gets its thermal energy initially from the at least 1800 molecules of

water which surround it, and it does not matter particularly if that 1800 molecules of water is actually 1799 molecules of water and another molecule of reactant. We have not, however, found the ingenuity to explain how these hypotheses may bring about a decrease of fraction removed with increase of initial concentration at low temperatures. Also, we can offer no explanation why the oxygen pressure has no effect on the rate or the extent of conversion, except that it may possibly be a spurious statement, since the data do not contain experiments at varying oxygen pressures. The solubility of oxygen in the liquor at reactor temperatures and pressures is very small compared to the amount required to oxidize COD in a gallon of liquor. It may be concluded therefore that oxygen transfer must occur during the course of oxidation.

One result for the above theoretical reasoning might well be emphasized to preclude a likely misconception concerning the process itself. In a conventional reaction in which the reactants are only in part converted to product to yield an intermediate fraction removed, one would expect that the product could be put through a second identical process, i.e., more time at the same temperature, to accomplish a successive increment of reaction. Indeed, this is suggested by some of the statements in the literature. For example, "the 80% COD reduction obtained in these runs can be improved, if such is desired, by going to higher temperatures and/or holding time." This seems to suggest that further holding time at the same temperature would increase the fraction removed. However, the experiments to which this statement refers were those of the West-Southwest pilot plant in which the reactor time was already 171 minutes, which is well out on the flat portion of the curve showing fraction removed versus reactor time. An improvement might be made by going to a higher temperature and a higher holding time, or even by a higher temperature and a lower holding time; but it is not correct that improvement could be achieved by going to a higher temperature or a higher holding time. Thus, putting the product material back through the reactor for a second pass would be ineffective, whereas putting a feed material of the same low COD concentration through the reactor (the first time) would be effective.

Plant Design: Data, Assumptions, Computations

From basic data in the referenced literature, information gained by personal visit, inquiry to certain manufacturers, and

some general chemical engineering process design knowhow, preliminary process and plant designs were made for nine cases of interest as follows:

3 concentrations:	organics	inorganics
	100,000 ppm	200,000 ppm
	3,000 ppm	6,000 ppm
	100 ppm	200 ppm
3 flow rates:	10,000,000 gpd	
	100,000 gpd	
	1,000 gpd	

BASIS FOR PLANT COST

In preparation for this the design of the Chicago plant as described in the literature was reconstructed in order to determine some of the bases used. Unit cost figures were determined or unit cost curves constructed especially for the major equipment items consisting of sludge pumps and motors, compressor and motor, heat exchangers, turbine, and fired heater. These were based on the data for designed wet oxidation plants and where possible supplemented by general process engineering data. Also, the ratio of principal items of equipment cost to total plant cost as a function of total cost of the former was plotted as a curve derived from plant design data in the references.

The reactor conditions were set to achieve 80 percent reduction of COD concentration, according to the design data in Chapter II. Since there are no experimental data for f when the amount of water evaporated is large, Cases 1, 4, and 7 were material balanced for $f_{corr} = 0.80$ with reactor design for $f = 0.80$. The shape of the reactor was chosen to provide as a maximum approximately the same rate of air feed per square foot of cross section, as was used in the Chicago reactors. In some cases the air flow was not concurrent with the liquor flow at these conditions. To simplify the computations, a standard air feed of 10 percent stoichiometric excess over that required for the 80 percent oxidation was used in all cases. This gave approximately 2 percent oxygen in the effluent gas. The reactors are stainless steel clad.

The air compressors were multi stage piston compressors.

The heat exchangers were designed to operate at the same average overall heat flux used in the Chicago plant, namely,

15,800 Btu per square foot per hour. The fuel fired heater was necessary for cases which did not produce sufficient heat to maintain the process. The heaters were of stainless tube construction.

In the estimates the cost of land is not included nor is that of off-site facilities such as collecting systems, disposal systems, etc.

One of the first developments in the process design revealed that the wet oxidation process could not be used on the highest concentration cases. For each pound of organics removed a certain amount of air was required for oxidation. This air carried water out of the reactor as saturated steam. The amount of water carried out could be adjusted within limits by adjusting the pressure and temperature in the separator. However, at high concentrations the amount of water carried out was enough that if the inorganic concentration was also high, the effluent was oversaturated in inorganics. It was assumed in the preliminary calculations that the inorganics were entirely NaCl or rather that they had a corresponding solubility of about 25 percent. For the preliminary assumptions it was found that an organic content of 60,000 ppm and an inorganic content of 120,000 ppm in the feed liquor was as high as could be tolerated without exceeding about 20 percent concentration of inorganics in the effluent. Accordingly, the concentration of the high concentration cases was dropped to these figures. The problem was worked out only for the highest rate of 10 mgd. A higher concentration could be tolerated at the intermediate and lowest rates, but for uniformity the initial concentrations were taken as the same in all three cases.

In computing the plant costs, the ratio of equipment costs to plant costs was applied to the sum of the major items of equipment. All costs used in this report have been adjusted to a Marshall and Stevens Index of Chemical Process Industries Equipment Costs of 240, which corresponds, approximately to the 1960 value of the Index (1926 = 100, 1913 = 59.0).

The pertinent design data on the principal items of equipment and the unit plant costs are detailed in Table 3.

BASIS FOR DISPOSAL COSTS

The following list shows the assumptions used in computing production costs.

Operation:	365 days a year

Utilities:	
Electric power consumed:	over 100,000 kw, 5 mills/kwh under 100,000 kw, 7 mills/kwh
Sale of byproduct power:	4 mills/kwh
Fuel:	25¢/10^6 Btu.
Direct Labor:	$2.00/hr.
General overhead, administration, and payroll extras:	45% of direct labor
Maintenance materials, labor, and operation supplies:	2% of total plant cost annually
Amortization:	7.4% of total plant cost annually (20-year plant life and 4% interest)
Taxes and insurance:	2% of total plant cost
Interest on operating capital:	4% of one-sixth of annual operating cost

The process design, being preliminary, was not subjected to economic optimization.

The disposal costs in dollars per 1000 gallons and in dollars per ton of organics are shown in Table 2.

DISPOSITION OF "ASH"

In the wet oxidation of sewage sludge there remain in the effluent suspended solids, mostly inorganic, which have been termed "ash" in reference to the ash from convential combustion. Typical sludge consists of 65 percent volatile matter equivalent to the organics and 35 percent non-volatile matter equivalent to the inorganics. Thus, for every ton of organic matter entering a sewage sludge wet oxidation process, there enters 0.539 ton of inorganics, some of which appear as suspended solids in effluent. The suspended solids are removed in sewage sludge oxidation by recycling to the conventional process. In wet oxidation for ultimate disposal of advanced treatment waste these solids would be settled separately, filtered, and disposed of in land fill. The possibility also exists that useful products, e.g. ceramic products, might be made out of them.

Table 3. PLANT DESIGN DATA FOR ULTIMATE DISPOSAL BY WET OXIDATION

	Case Number								
	1	2	3	4	5	6	7	8	9
Ambient feed liquor									
Organic content, ppm	60,000	3,000	100	60,000	3,000	100	60,000	3,000	100
Inorganic content, ppm	120,000	6,000	200	120,000	6,000	200	120,000	6,000	200
Total content, ppm	180,000	9,000	300	180,000	9,000	300	180,000	9,000	300
COD, (x_0), g/l	120	6	0.2	120	6	0.2	120	6	0.2
1000 gallons/day	10,000	10,000	10,000	100	100	100	1	1	1
gpm	6,950	6,950	6,950	69.5	69.5	69.5	0.70	0.70	0.70
organics, tons/day	2,820	126	4.16	28.2	1.26	0.0416	0.282	.0126	0.000416
Density, g/cm^3	1.13	1.01	1.00	1.13	1.01	1.00	1.13	1.01	1.00
Reactor conditions									
Temperature, °F	541	547	550	503	528	531	503	528	531
Time, (t_R), min	25	50	50	108	108	108	108	108	108
Pressure, psig	1,500	1,500	1,500	1,000	1,000	1,000	1,000	1,000	1,000
Effluent liquor									
Organic content, ppm	20,900	602	20	20,600	602	20	15,200	603	20
Inorganic dissolved solids, ppm	214,000	6,020	200	206,000	6,020	200	152,000	6,030	200
COD, g/l	41.8	1.2	0.04	41.2	1.2	0.04	30.4	1.2	0.04
1000 gallons/day	5,280	9,960	9,998	55	99.6	99.9	0.735	0.995	0.999
Organics, tons/day	564	25.2	0.83	5.6	0.25	0.0083	0.056	0.0025	0.000083
Main equipment parameters									
Unit reactor volume, gal	15,000	15,000	15,000	7,500	7,500	7,500	75	75	75
Number of units	12	24	24	1	1	1	1	1	1
Unit cost									
$/gal	12	10	10	9.30	9.30	9.30	13.33	13.33	13.33
% of main equipment	5.8	16.5	17.5	12.4	20.2	21.6	5.9	4.6	5.1
Air compressor including driver, receiver, and separator, unit bhp	12,000	1,810	29	1,300	73	1	15.7	1	use
Number of Units	12	4	1	1	1	1	1	1	bottled
Unit cost									oxygen
$/bph	104	130	172	119	207	3,000	312	3,000	
% of main equipment	43.4	4.6		27.4	4.4	0.9	28.8	13.8	
Sludge pumps, unit bhp	175	175	175	29	29	29	0.5	0.5	0.5
Number of units	36	36	36	2	2	2	1	1	1
Unit cost									
$/bph	700	700	700	850	850	850	3,000	3,000	3,000
% of main equipment	12.7	21.3	22.6	8.8	14.3	15.2	8.8	6.8	7.7
Heat exchangers, unit ft^2	1,000	1,000	1,000	100	100	100	64	96	96
Number of units(approximate)	56	95	95	5	9	9	1	1	1
Unit cost									
$/ft^2	112	112	112	180	180	180	150	150	150
% of main equipment	18.0	51.9	54.5	15.8	46.9	50.1	56.5	65.7	71.3
Expansion engine and generator, unit hp	71,600	6,200		2,040	72				
Number of units	3	1	none	1	1	none	none	none	none
Unit cost									
$/hp	32.30	87		98	196				
% of main equipment	20.1	2.6		35.6	4.1				
Net electric power, (credit), kw	(40,000)	7,330	6,180	(390)	76	63	17.5	6.5	5.75
Fired heater, 10^6 Btu/hr absorbed		32	55		1.5	2.23		0.029	0.036
Number of units	none	4	4	none	1	1	none	1	1
Unit cost									
$/$10^6$ Btu/hr		5,000	4,780		23,400	17,700		70,000	70,000
% of main equipment		3.1	5.4		10.1	12.2		9.1	12.9
Fuel, 10^6 Btu/hr		152	260		1.76	2.62		0.042	0.052
Ratio of equipment cost to plant cost	0.57	0.57	0.58	0.37	0.34	0.34	0.25	0.25	0.25
Labor, man per shift [a]	6	4	4	1	1	1	0.1	0.1	0.1
Cooling water, 1000 gal/day	167,000	1,050	3.9	1,640	20	neg	16	neg	neg
Evaporated water condensed, 1000 gal/day	4,300	27	0.1	42	0.5	0.1	0.265	0.005	0.001
Costs									
Unit plant cost, $/gpd	6.12	3.62	3.35	15.20	10.10	9.50	69.00	97.50	77.50
Unit disposal cost, $/1000 gal	1.59	1.38	1.34	5.10	4.10	3.94	31.60	35.90	32.80
Unit disposal cost, $/ton of organics removed	7.02	137	4,000	22.60	406	11,800	140	3,560	98,500
Daily disposal cost, $/day	15,900	13,800	13,350	510	410	394	31.60	35.90	32.80
Daily fixed costs, % of disposal costs	121	81.9	78	93.2	77.0	75.4	68	76.2	74.0

[a] Reference 3

GPO 805-035-3

In the present study, which was applied essentially to secondary sewage treatment effluent, there are no data indicating the amount of suspended solids to be expected in the effluent since the original feed consists largely of dissolved organics rather than suspended organics. Accordingly, the disposition of the effluent suspended solids has not been considered in this study. The effect would be small however. Preliminary estimates of the cost of disposal by wet oxidation of the 2 percent suspended solids in the effluent for Case 1 show plant costs of $0.07 per gpd and disposal costs of $0.09 per 1000 gallons, which includes the cost of hauling 10 miles.

Correlation of Costs

Figures 5 and 6 show the correlation of unit plant costs and unit production costs for the nine cases in Table 2 according to the liquor feed rate and the initial COD concentration. The plant costs vary from $3 to $6 per gpd for the 10-mgd plants to $69 to $87 per gpd for the 1000-gpd plants. The established design costs for the Wausau, Wheeling, and Chicago plants and the package plant correlate satisfactorily with these computations (see Table 4). The unit production costs vary from about $1.40 per 1000 gallons for the 10 mgd plant to $36 per 1000 gallons for the 1000-gpd plants.

Table 4. LOADING COSTS

Case No.	1	2	3	4	5	6	7	8	9
1000 gpd	10,000	10,000	10,000	100	100	100	1	1	1
Contaminants, ppm	300,000	9,000	300	300,000	9,000	300	300,000	9,000	300
Tons/day inorganics	9,660	250	8.32	96.6	2.50	0.0832	0.966	0.0250	0.000832
Time between 40-ton shipments, days	0	0	4.8	0	16		41.5		
years						1.32		4.4	132
Capital costs, loading and storage, and equipment, $			10,000		10,000	10,000	10,000		
Fixed charges, $/1000 gallons			0.002		0.027	0.027	2.67		
Drum cost, $/1000 gallons								0.25	0.00832
LCL freight factor	1	1	1	1	1	1	1	4	4
Handling cost, $/1000 gallons	0.181	0.0047	0.000016	0.181	0.0047	0.000016	0.181	0.0047	0.000016
Total loading cost, $/1000 gallons	0.181	0.0047	0.002	0.181	0.032	0.027	2.85	0.255	0.00834

In later stages of this series of studies the absolute magnitude of the figures in Table 4 will be compared with those for competitive ultimate disposal processes. Within the scope of the present study, however, it may be concluded that feed rate affects unit cost considerably more than feed concentration. For example, the extreme range of feed concentration is 600-fold.

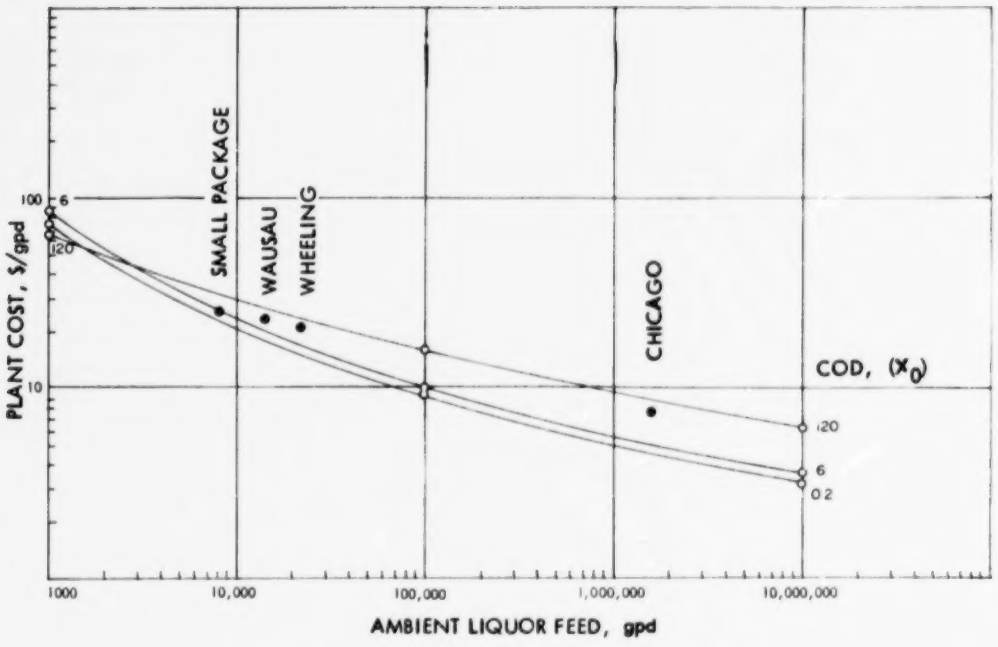

Figure 5. Plant cost for wet oxidation disposal.

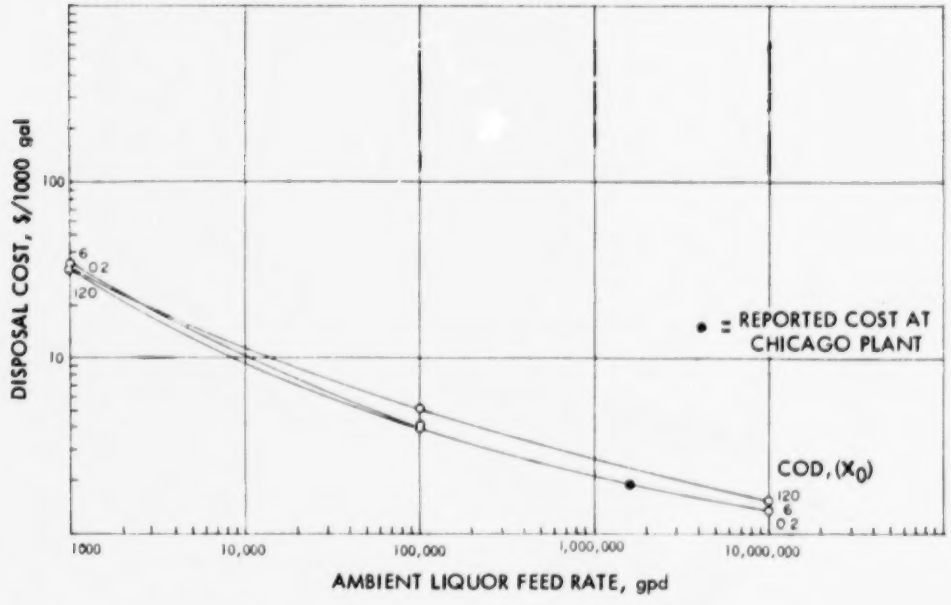

Figure 6. Disposal cost by wet oxidation.

At the greatest, for the 100,000-gpd plants, an increase of feed concentration by 600-fold increases the unit plant cost to 1.6 and the unit disposal cost to 1.3 times that at the lowest concentration. However, a 600-fold decrease in feed liquor rate increases the plant cost to 5-15 times and the production cost to 6-14 times that of the higher feed rate, depending upon the position in the feed rate scale. Stated in another way and generalized, the

600-fold range in concentration has only as much effect as a two- to five-fold range in feed rate. Furthermore, it should be noted that a decrease in feed rate increases the cost and an increase in feed concentration also increases the cost.

Figure 7 provides the basis for exploring the logistical implications of the above generalizations as they apply to the relation between advanced treatment processes and ultimate disposal by wet oxidation. In this Figure the data are arranged so that disposal cost in dollars per ton of organics oxidized can be plotted against organics in the feed in tons per day.

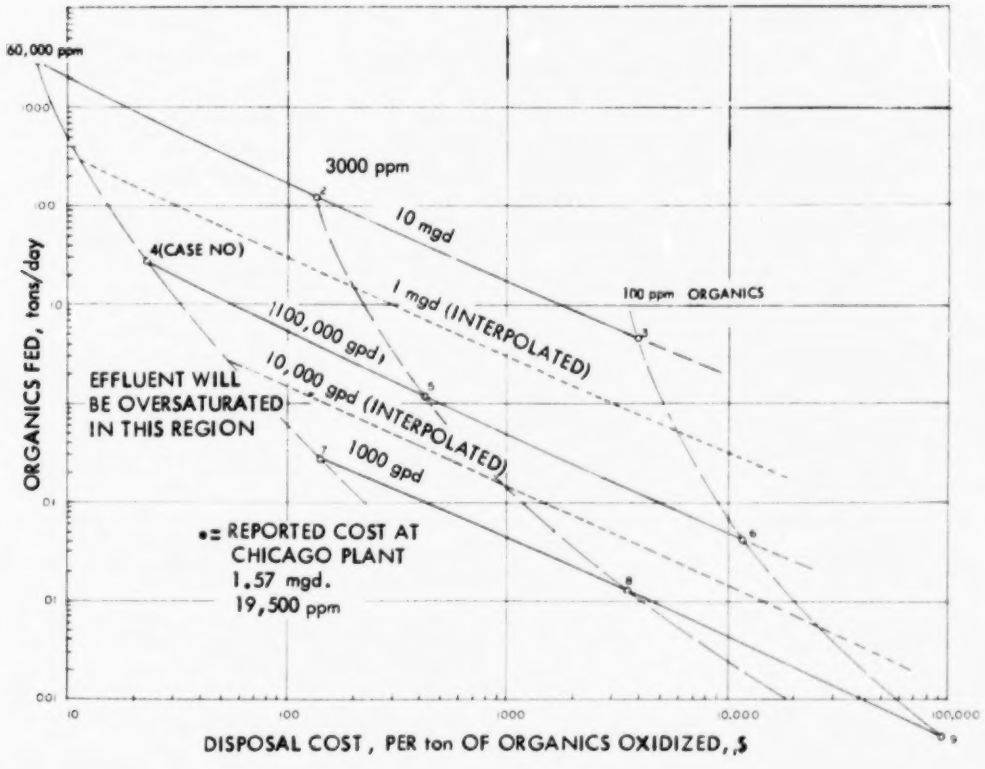

Figure 7. Per-ton disposal cost by wet oxidation.

The approximately straight lines represent this relationship for constant feed rates. A third set of coordinates is superimposed in the curved dashed lines representing constant initial concentration of organics. The designer of an advanced treatment plant which is not in itself an ultimate disposal plant may choose between producing a large quantity of low concentration wastes or a small quantity of high concentration wastes. The question to be explored is which of these alternatives produces the lower cost for ultimate disposal. (Of course the ultimate question is which

of the alternatives produces the lowest cost for the sum of advanced treatment and ultimate disposal, but development of that optimum will have to await the development of relationships similar to these for the advanced waste treatment processes themselves.) The fixed point in advanced treatment plant operation is the total tons per day of organics which must ultimately be disposed of. This comprises the vertical axis in Figure 7. The Figure shows that for any fixed tons per day of organics wet oxidation is cheaper if this fixed tonnage is contained in the smallest possible amount of feed liquor. This means, and contrary to what might be expected from first glance at Figure 6, that ultimate disposal is cheaper when a small quantity of a concentrated waste is handled, than when the same amount of organics is handled as a large quantity of dilute wastes. This arises because the slope of the curve in Figure 6 is less than 45 degrees. Also to be noted in Figure 7 is the exclusion area to the left of the 60,000- ppm coordinate. Costs lower than this cannot be achieved because concentrations higher than this would result in oversaturation for the standard quantity of associated inorganics. Of course, if the ratio of inorganics to organics is reduced then one could go to higher initial concentrations.

The crux of the discussion is apparent in Figure 8 which shows the disposal cost in dollars per day for various tons per day of organics fed. The nearly vertical solid lines are lines of constant feed rate from which it can be seen that the daily disposal cost is largely a function of feed rate and is influenced very little by concentration, that is by tons per day of organics. The dashed coordinates represent lines of constant initial concentration from which it may be seen that at any fixed tons per day of organics the cheapest disposal is that for the most concentrated feed. Again the area to the left of the 60,000 ppm coordinate is an excluded area in which the effluent would be oversaturated with inorganics, and therefore, the 60,000 ppm coordinate represents the minimum daily disposal costs that can be achieved for any given tons per day of organics.

The main purpose of developing these figures was to compare them with other methods of ultimate disposal and to optimize the advanced-treatment-ultimate-disposal overall process. Since data for neither of these two comparisons have yet been developed, such use cannot be made of data in this report. However, as an example one may explore the relationship between wet oxidation and conventional sewage treatment. For a town of 60,000 population, the sewage flow would be about 10 mgd containing about 5 tons per

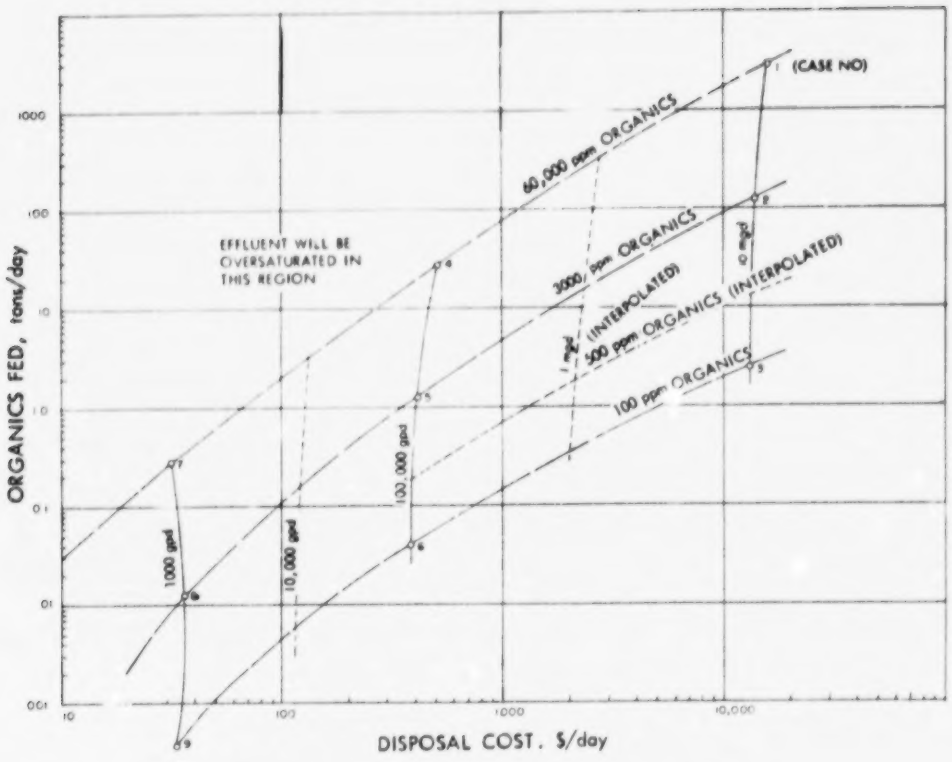

Figure 8. Daily disposal cost by wet oxidation.

day of organics at a concentration of about 125 ppm. To give this sewage primary and secondary treatment, reducing the organics to 50 ppm, would cost roughly $700 per day.* Wet oxidation disposal of the 4 tons per day of organics in the primary and secondary sludge, containing say 5 percent dry matter, would cost an additional $200 per day, according to Figure 8. This disposal process would, however, replace the present sludge digestion and disposal processes included in the $700 per day costs. The direct use of wet oxidation on the raw sewage from the 60,000 population would contain 5.25 tons per day of organics at 125 parts

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* "Annual operation and maintenance costs" would be about $1.00 per capita annually.[12] It appears that this figure does not include the fixed charges on the investment, which for purposes of calculation here will be taken[13] at 3.40 1913 (ENR) dollars per capita capital investment. Converting to ENR = 825 for 1960, one obtains a fixed cost at 9.4 percent annually of $435 a day, which added to the operation and maintenance costs of $164 a day provides roughly $700 a day or $70 per million gallons for conventional treatment costs.

per million and would cost about $13,000 per day, according to Figure 8. Finally, one may inquire what wet oxidation of the secondary effluent from its original 50 ppm organics to an effluent of 10 ppm organics might cost. The organics in the secondary effluent at 50 ppm would amount to about 2.1 tons per day, and the daily disposal cost, from Figure 8, would be about $13,000 per day, also.

These questions have been explored as illustrations of the types of logistical analyses that can be made and also to answer the quite obvious question: "Why not use wet oxidation directly on raw sewage, on primary effluent, or on secondary effluent?"

It is noted that the fixed costs constitute a very large proportion of the total disposal costs. This means that the daily disposal cost would be rather insensitive to actual throughput rate. If the plant operates at more than rated capacity, the daily operating cost would not increase much. On the other hand, as is more likely, if the plant operates at only a fraction of its capacity, the daily operating costs would not decrease very much.

An example of how the data here developed may be used in research guidance and project evaluation follows. Taking the same population and flow conditions let it be assumed that the organics in the secondary effluent are 100 ppm. From Figure 8 the costs of wet oxidation ultimate disposal may then be developed as a function of the percent recovery of product water in the advanced treatment process. Table 5 shows the developed data.

Table 5. UNIT COSTS OF WET OXIDATION DISPOSAL OF ADVANCED TREATMENT WASTES FROM A CITY OF 60,000 POPULATION WITH A WASTE FLOW OF 10 mgd (100 ppm ORGANICS IN SECONDARY EFFLUENT)

Water recovered, %	Concentration of organics in waste, ppm	Cost of water produced $/1,000,000 gal
99.83	60,000	15.5
96.67	3,000	91.1
80.00	500	538.0

It is apparent that for any reasonable ultimate disposal cost it would be necessary to run the advanced treatment process to very high recoveries. This had been indicated qualitatively in earlier

reports. The present study provides the quantitative figures for wet oxidation. It must be remembered that in this example additional ultimate disposal costs must also be incurred for disposal of the inorganics remaining in the waste. The graphs provided in the present study allow such logistical explorations over the wide range of parameters chosen.

Areas for Further Investigation

A number of areas beyond this preliminary study remain for further investigation.

The major uncertainity concerning the process design itself lies in the extrapolation of the conversion kinetics from the lowest experimental value of about 13 grams per liter COD to extreme dilution. The amount of pilot plant design data available in the sources quoted in this report is very scanty. Both of these deficiencies should be corrected by further experimental investigation preferably under continous conditions, although some information can be gained from batch experiments, especially concerning the ultimate degree of oxidation and the end products.

It is not unlikely that some fundamental investigations into the kinetics of the wet oxidation reaction would yield some useful results. For example, it would be well to investigate the influence of oxygen pressure over a wider range of variables than appears in the present evidence. It is shown in the body of the report that oxygen transfer must occur, and if this is the controlling factor in the reaction rate, ways might be devised to speed up this process. Furthermore, inquiries into the mechanisms of the reaction may show how to achieve the immediate breakup into small molecules which is postulated as one of the hypotheses.

There are substances which act as oxidation catalysts, and an investigation should be undertaken to determine whether any practical catalysts can be developed, not so much to speed up the reaction as to reduce the temperature at which high conversions may be achieved. A lower temperature would allow operation at a lower pressure and reduce the over-all costs.

Experimental studies should be undertaken on the tolerance of the reactor and the rest of the system for relatively small amounts of precipitated inorganic solids resulting from oversaturation. In the body of the report it has been assumed that the inorganic solids are similar to NaCl, i.e., that they have a relatively high and narrow range of solubilities. However, the inorganic content of actual wastes would contain a spectrum of

solubilities which for the present purposes may be thought of rather as a spectrum of "percent of solubility". (Example: If the water evaporated reduces the liquid volume to one-half of its original value, then there would be precipitated all materials whose "percent of solubility" was greater than 50 in the original liquor.) The effluent would have already contained suspended solids and presumably a small addition to the total suspended solids could have been handled in the normal equipment. The clarified effluent would then be saturated with respect to these particular solids.

The present report takes a constant ratio of inorganics to organics and arrives at a limiting value of the organic content beyond which enough water would be removed to precipitate some of the inorganics. Process design and cost analysis should be run through with differing ratios of inorganics to organics to determine the effect on this limiting line of changes in that ratio.

The present process design has been very preliminary in nature with little opportunity for optimization. The next stage of accuracy in process design and cost estimation should involve optimization of the design to balance the various operations for the lowest overall costs.

The study was undertaken with the idea that wet oxidation can be controlled by varying the time-temperature regimen, that practically 100 percent oxidation could be acheived if time or temperature were sufficient. It was discovered in the course of the study that this is not the case and that at practical operating conditions residual organics remain in the effluent. Left unsolved therefore is the problem of ultimate ultimate disposal of this residuum. In the recommendation on kinetic and mechanism studies it was indicated that some attention should be given to achieving the highest possible breakup of the initial molecules into CO_2, with a minimum of low-molecular-weight organics such as acetic acid. This is one approach to the problem. A second approach would be to determine what supplementary processes might be applied to reduce to zero these residual organics from wet oxidation. If the residual material is acetic acid one should seek for insoluble acetates. Most inorganic acetates are soluble but basic ferric acetate and basic aluminum acetate are insoluble in water. Studies should be undertaken to determine whether these or other compounds are suitable for reducing the low-molecular-weight residual materials from wet oxidation.

Reduction of organic residuals may be accomplished by recycling the wet oxidation effluent to the conventional secondary

sewage treatment process. This was done in the Chicago pilot plant and will be done in the Chicago plant, since the low-molecular-weight residual organics from wet oxidation are in turn very amenable to biological oxidation. This constitutes an ideal disposal method for these residuals from wet oxidation of sewage sludge. Further thought reveals its failing as a method for advanced treatment, however. The difficulty lies with the inorganics in the waste. In sewage treatment the entering inorganics pass through the process and into the secondary effluent. A small portion of the liquor containing the sludge is cycled through the wet oxidation process, and the residual liquors from that are recycled to the secondary treatment. The inorganics found in the sludge and the sludge liquor are inappreciable in the total inorganics and are bled off in the secondary effluent. With advanced treatment, however, there is no place to bleed off the inorganics, since the products from the advanced treatment plant will be pure water and the waste, which is the feed liquor to our study. This feed liquor contains all the inorganics in the original water plus all those added by use. It cannot be recycled to the secondary stage of the conventional treatment plant, preceding the advanced treatment, because the inorganics would soon build up in the recycled liquor and one would be operating the entire treatment complex as a saturated inorganic slurry. Some further analysis of this possiblity is worthwhile and should be undertaken. To overcome such difficulties, one would pass the wet oxidation effluent to a separate secondary treatment which might even so be operating, at high salt content, depending on the degree of recovery of water in the advanced treatment stage. At any rate, by this sequence practically all the organics would have been removed, leaving only the inorganics in the residual liquor. Further studies are obviously needed on the disposal of these inorganics, but this is beyond the scope of the present project.

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Logbook References

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ULTIMATE DISPOSAL
BY
INCINERATION

Part 2.

INCINERATION

Summary

In some of the processes for the production of water being investigated under the Advanced Waste Treatment Research Program of the U.S. Public Health Service, water is extracted from the raw material or contaminants are removed therefrom to yield the ultimate product and a waste stream consisting of the contaminants and possibly some of the reagents in a more or less concentrated form. A preliminary process design and cost estimate was made for ultimate disposal of these liquid wastes by incineration to substantially organic-free dry solids followed by hauling to the sea coast and dumping offshore. The process parameters were concentration of contaminants and daily volume taken respectively as 300,000, 9,000, and 300 ppm total organic and inorganic contaminants and 1,000, 100,000 and 10,000,000 gpd. Parameters introduced by the hauling situation were the cost of hauling waste solids to the sea coast from various points in the nation represented by contour lines of equal cost for hauling waste solids to the sea coast in dollars per ton.

Consideration was given to incineration by Atomized Suspension Technique, a box furnace, flash dryer - Nichols Herreshoff furnace combination, and three processes under development for calcination of nuclear wastes. Although an exhaustive analysis was not made, the preliminary indications were that only the flash dryer - Nichols Herreshoff apparatus could be chosen as actually operable. Investment and operating costs were estimated for this design over the range from 412 to 4,120,000 pounds per hour of 30 percent solids feed corresponding to 1000 gpd to 10 mgd of advanced treatment wastes. Unit costs in $ / 1000 gallons of incinerator feed ranged from $5.08 at 10 mgd to $55.10at 1000 gpd; the corresponding unit investment cost per gallon per day of incinerator feed ranged from $9.62 to $97.80 respectively.

Since concentration in an evaporator was judged, and later proven, less expensive than evaporation in the incinerator, evaporators were designed to bring all feeds to 30 percent solids. In the smaller capabilities, this was done in one step with a

single-effect evaporator. However, at the higher capacities, the major portion of this pre-evaporation, to 10 percent contaminants, was very similar to saline water conversion evaporation, and accordingly, saline water distillation cost estimates were analyzed, adjusted for uniformity, and used as such. Single and triple-effect evaporators were designed and the unit costs per 1000 gallons of water evaporated were correlated according to capability in gpd of water evaporated, along with similar data for saline water conversion by distillation and for evaporation of water in the flash dryer - Nichols Herreshoff apparatus. The results were presented as curves over the range 1000 gpd to 80 mgd.

Class rail rates for hauling waste solids were determined over sample routes in the western, midcontinent, and eastern sections of the country and correlated on the basis of straight line distance from the nearest ocean. A map was generated to show lines of constant cost in dollars per ton for hauling waste solids to the nearest ocean, the Pacific, the Gulf, or the Atlantic. The nature and costs of loading the waste solids into hopper cars for rail hauling were investigated.

The possibilities of decreasing the hauling cost by application of commodity rates rather than class rates, and by utilizing rail and barge hauling were briefly investigated. A generalized pattern of hauling in the ice-bound and the ice-free seasons was developed in a preliminary way.

Costs were estimated for hauling the wastes 30 miles to sea and dumping them.

To a subtotal cost for pre-evaporation, incineration, loading and storage, and ocean dumping was added a rail hauling cost in dollars per ton of solids. At approximate U.S. average conditions of $10 per ton for hauling, the unit cost of advanced treatment waste in dollars per 1000 gallons varied from $0.565 for 10 mgd and 300 ppm to $68 for 1000 gpd and 300,000 ppm. Unit costs increased with decreasing capability and with increasing concentration. The cost of hauling is negligible except at the high concentrations.

Approximate unit investment costs, which could not be estimated accurately because of the method of including already developed saline water conversion costs, ranged from $1.53 per gpd at 10 mgd and 300 ppm to $108 per gpd at 1000 gpd and 300,000 ppm. The trend was the same as for operating costs, decreasing with decreasing concentration and increasing with decreasing capability.

Unit disposal costs by incineration with no hauling of the waste solids at 180,000 ppm were compared with the corresponding

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unit disposal costs for wet oxidation. For the latter, no cost was incurred for disposing of the waste liquors. Incineration costs under these circumstances are above wet oxidation costs at all capabilities.

Ultimate disposal of 10,000 to 100,000 gpd by injection would cost the same as incineration of 1000 gpd under average U.S. conditions. Likewise 1 to 10 mgd could be disposed of by injection for the same daily costs as 100,000 gpd by incineration.

As an example for the guidance of advanced waste treatment research, there was compared the daily cost of incineration for advanced treatment processes for a community of 60,000 population producing 10 mgd of a 300-ppm feed to the advanced treatment process. The cost of disposal by incineration per million gallons of water produced in the advanced treatment plant would be $670 at 50 percent recovery in the advanced treatment process, and $24 at 99.9 percent recovery, which would produce a waste of the upper limiting concentration. However, the pre-evaporation stages of the incineration process also produce distilled water which, if potable, could be added to the total production. In terms of this total water produced for the above example, the disposal cost is $335 per million gallons at 50 percent recovery and $24 at 99.9 percent recovery. These latter costs are of the same general magnitude as the corresponding costs for ultimate disposal by wet oxidation and are considerably higher than the corresponding disposal costs by injection. All of the above comparisons are made under conditions of average values of the pertinent parameters.

A number of areas for improving the present study are suggested. Especially important are studies of transportation economics, of pre-evaporation costs and condensate quality, and of oceanography.

Pre-Evaporation

In the early stages of the study it was anticipated that the Atomized Suspension Technique (AST) would be used for converting the feed liquors directly to incinerated dry solids. However, for reasons detailed in Chapter III the flash dryer - Nichols Herreshoff furnace incineration system was chosen. This equipment is essentially for drying rather than evaporating and requires a feed liquor of rather high solids content. It was hypothesized that it would be cheaper to evaporate the bulk of the water from the low-concentration feeds by means of evapora-

tors. However, it was thought that 30 percent solids, that is the highest feed concentration, was about as high as one would be able to carry efficient evaporation in an evaporator. Accordingly all the dilute feeds were pre-evaporated to 30 percent.

EVAPORATION TO 10 PERCENT

A great deal of work and estimating has been done on saline water conversion in which a feed of 3.5 percent solids is taken to 7, 10, or slightly higher percent solids. The estimated costs for doing this at the high capabilities at which the estimates are customarily made are considerably lower than costs estimated for conventional evaporation at the lower capabilities at which such estimates are usually made. Accordingly, in this preliminary study it seemed proper to utilize for pre-evaporation to 10 per cent the already developed estimates for saline water conversion by distillation. From among these, estimates showing the lowest cost and not tied to power production were chosen and were adjusted to achieve a greater degree of uniformity by computing for 4 percent interest, 2 percent taxes and insurance, and approximately 20-year amortization, and by adjusting to 1961 conditions by the Engineering News-Record Construction Cost Index. The estimates chosen were

Badger	1955	10-effect LTV vapor compression (steel)[3, 7]
Dodge	1958	single-effect, forced circulation vapor compression[4]
Fluor	1959	52-and 42-stage multiflash nuclear steam[5, 8]
Fluor	1957	7-stage multiflash LTV combination nuclear steam[6]

The data are shown in Figure 1. It is to be noted that the 10-effect LTV vapor compression plant is estimated for steel construction. It is not certain that steel will be an adequate construction material. The 7-stage multiflash plant estimated with steel construction shows a water cost (at highest capability) about 20 percent below the cost with the alloy plant. Thus, as a first approximation the 10-effect plant steel costs might be increased by 25 percent to conform to alloy construction.

BASIS FOR EVAPORATION DESIGN

The product from the first evaporation to 10 percent contaminants becomes the feed for the second preliminary evaporation from 10 percent to 30 percent. Preliminary designs for the required evaporators were based on conventional chemical engineering design factors, specifically the following:

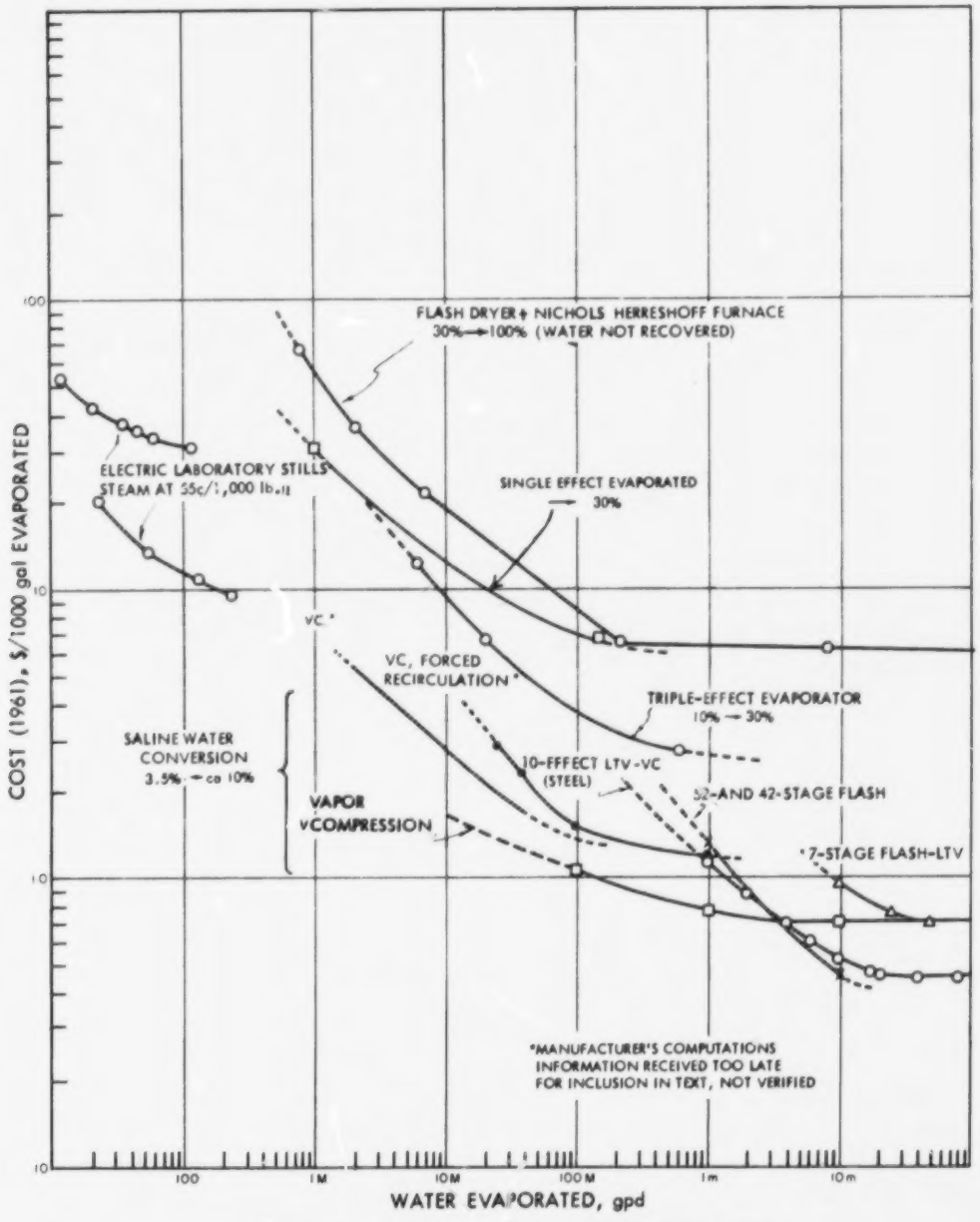

Figure 1. Cost of evaporating water from salines.

Evaporator construction material: stainless steel

Operation: 330 days per year

Electric power: Under 100,000 kw, 7 mills/kwh

Cooling water: 40° F rise

Fuel: 25¢/1,000,000 Btu

Steam: 55¢/1,000 lb

Direct labor: $2/hr

General overhead, administration and payroll extras: 45% of direct labor

Maintenance materials, labor, and operating supplies: 2% of total plant cost annually

Amortization: 7.4% of total plant cost annually (20-year plant life and 4% interest)

Taxes and insurance: 2% of total plant cost

Interest on operating capital: 4% on 1/6 of annual operating cost

Cost index: All costs adjusted to Marshall and Stevens Index of Chemical Process Industries Equipment Costs of 240, which corresponds[9] approximately to the 1961 value of the index (1926 = 100, 1913 = 59)

The data and results are shown in Table 1 and are plotted in Figure 1. Cases 1, 4, and 7 do not require evaporation since the feed is already at 30 percent. In Case 6, the feed for the second pre-evaporation is only 280 gallons per day and this was considered too small for cost estimation.

Table 1. EVAPORATOR DESIGN DATA

Used in case no.	2	2	3	5	8 and 9
Waste, 1000 gpd	10,000	10,000	10,000	100	1
Water evaporated, 1000 gpd	602	602	20.2	6.04	1.0
No. of effects	1	3	3	3	1
No. of units	4	1	1	1	1
Capacity each unit, 1000 gpd evaporated	150	602	20.2	6.04	1.0
Feed concentration, ppm	100,000	100,000	100,000	100,000	low
Concentrate, ppm	300,000	300,000	300,000	300,000	300,000
Capital cost, $	1,429,000	1,478,000	100,600	64,000	17,800
Labor, men / shift	1	1	0.75	0.5	0.25
Labor cost, $/365 days	25,400	25,400	19,000	12,700	6,350
Steam consumption, lb /lb H_2O evaporated	1.22	0.397	0.397	0.397	1.18
Steam cost, $/330 days	1,100,000	362,000	12,100	13,620	1,780
Cooling water, 1000 gpd	14,720	4,940	166	49	24.6
Cooling water, 1000 gallons/1000 gallons of water evaporated	24.5	8.2	8.2	8.1	24.6
Power, kw	380	190	33	11	2
Power cost, $/330 days	21,000	10,500	1,490	615	110
Fixed charges, $/365 days	163,000	168,000	11,550	7,300	2,030
Operating capital cost, $/365 days	8,800	3,800	295	162	68
Total annual cost, $/year	1,318,000	569,700	44,430	24,400	10,340
Unit cost, $/1000 gallons evaporated	6.65	2.77	6.64	12.25	31.3
Unit capital cost, $/gpd evaporated	2.37	2.45	4.98	10.60	17.8

Also plotted in Figure 1 are the data for incineration considered as an evaporation operation. They are shown here in order to confirm the original hypothesis that it is cheaper to evaporate water in standard evaporators than in the flash dryer - Nichols Herreshoff system.

It is noted that the cooling water requirement is shown in 1000 gallons per 1000 gallons of waste and is rather high, for example, 0.5 for Case 2 triple-effect evaporation. (See Table 5) To this in the overall process should be added the unit cooling water requirement of the saline water conversion type of evaporation to 10 percent contaminants; however, this figure was buried in the original data and therefore was not extracted for the

report. For a plant with adequate cooling water supply, for example at the sea coast or on a major river, the problem is not of great importance since the costs associated with pumping the cooling water have already been included in the estimated operating costs. For a plant located in the arid west, however, it would presumably be difficult to obtain feed water, much less cooling water. Accordingly, some charge would have to be made for the cooling water, considered as a material, as well as for pumping. Under such conditions one would substitute air cooling for water cooling at some increase in unit cost.

Incineration

ATOMIZED SUSPENSION TECHNIQUE

The Atomized Suspension Technique (AST) as developed by the Pulp and Paper Research Institute of Canada would seem ideal for converting advanced treatment wastes into an incinerated dry solid for disposal. Accordingly, a special visit was made to the Institute, the Atomized Suspension equipment inspected, and data obtained on its operating characteristics. In addition, a large amount of published literature, not referenced here, was obtained on the process.

Despite this preparation the Atomized Suspension Technique was not estimated for two reasons. First it was learned that the technique had been tried as a means for saline water conversion on concentrated brines and on inland brackish waters.[11] This work indicated that, unlike sewage sludges and other organic-containing liquors, saline brines in the AST process rapidly form an adherent deposit on the walls of the reactor. A major research investigation was envisaged to seek methods of overcoming this difficulty, but until the research had been undertaken one could not be sure that the obstacle could be overcome.

The second reason for not estimating on AST is that it does not appear, based on preliminary calculations, to offer any substantial advantage over the process used in this report, either in operating economy or in capital cost per unit of organic oxidation capacity or per unit of water evaporation capacity. This conclusion reflects the particular conditions of the advanced waste treatment incineration problem dealt with in this report.

BOX FURNACE PROCESS

It was also originally conceived that the advanced treatment waste liquid might be sprayed into a conventional box furnace, firebrick lined, which would be fired by oil, gas, or powdered coal, the water being evaporated and the organics incinerated in a single chamber. Although this would be the simplest type of equipment in construction, there are several problems inherent in the design of such a unit. These problems appear to require resolution before this method can be realistically proposed:

1. The problem of insuring that solid residues do not accumulate in the furnace either as particulate matter or as adherent deposit on the brick walls.
2. The temperature and residence time required to assure complete oxidation of the organic contaminants.
3. The cleanup facilities required to prevent atmospheric pollution by the stack gases.

As in the case of AST, the box furnace may have the potential for lowering the capital cost of an incinerator installation, but does not offer substantial improvement in thermal efficiency over the design used in this study.

FUSED SALT CALCINATION

A number of processes have been suggested and studied in the atomic energy work for reduction of processing wastes to dry solids by calcination. One of these involved calcination to a fused salt.[12,13] In this process fluxing agents naturally present or added to the mixture keep the melting point down to about 350° C. Costs were estimated at $920 to $670 per 1000 gallons, depending on the type of operation. It should be recognized that estimates for nuclear waste processing can be expected to be much higher than for incineration of advanced treatment wastes because in the nuclear work extensive precautions are necessary against radioactivity and against radioactive contamination of the off gases.

This process was not considered further for the present uses because of the difficulties of handling fused salts of the types to be expected in advanced treatment wastes and because oxidation would probably be slow in the fused mixture.

ROTARY BALL KILN

A second type of apparatus under development for nuclear waste processing is the rotary ball kiln in which the waste is fed to rotary kilns containing balls which act to continuously break up the solids formed and prevent scale buildup.[14, 15, 16] Temperatures of 700°C are reached. A 4-inch-diameter unit has been operated, and a preliminary design has been drawn for a 2.5-foot-diameter unit. Preliminary costs have been forecast at $863 per 1000 gallons (1959) for an approximately 4000-gpd unit.[17] Very little design information was available in the published literature on this process, and for that reason it was not estimated in this study.

FLUIDIZED-BED CALCINATION

Another method under study for nuclear waste processing consists of high-temperature calcination. Liquid waste is sprayed on a bed of fluidized particles maintained at about 500°C.[17, 18, 19, 20] A 12-inch unit has been operated, and a 60-gallon per hour demonstration unit was to go into operation January 1961. A conceptual design of a 7-foot-diameter, 25-foot-high calciner has been made. Costs for an approximately 4000 gpd plant have been estimated at $582 per 1000 gallons. As with the rotary ball calciner, it was believed that not enough performance or design data were available to allow an estimate in this study.

FLASH DRYER AND NICHOLS HERRESHOFF DESIGN

It was conceded that available conventional equipment was capable of drying and incinerating the materials. This was the flash dryer and Nichols Herreshoff furnace combination shown in Figure 2.

Feed liquor containing 30 percent solids is continuously mixed with recirculated dry solids to make a moist solid which can be easily handled as feed to a flash dryer system. For purposes of calculation herein, the water content of the flash dryer product material is estimated to be about 10 percent. This figure may vary considerably, but is not critical in the design. The drying is accomplished by hot gases from the Herreshoff furnace, and the solids are separated in an overhead separator for recycling. Some of the 10 percent moisture material is bled off from the cycle to feed the top hearth of the Herreshoff furnace. The furnace is heated with fuel; practically all of the water is

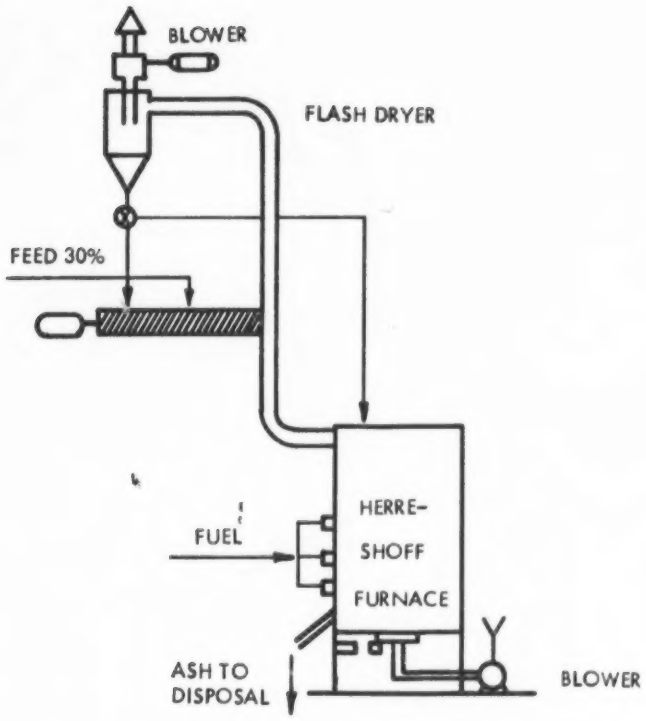

Figure 2. Schematic of incinerator system.

evaporated countercurrent to the hot gases on the top hearths; and incineration of the organics occurs on the lower hearths. The entire apparatus is termed the incinerator and is considered to have two parts, the flash dryer and the furnace.

Design and cost data were obtained directly from Nichols Engineering and Research Corp. by means of a vi letters, and telephone conversations.[21]

The costs of the Nichols Herreshoff furnaces were taken directly from the Nichols Engineering and Research Corp. data where possible. In some cases it was necessary to interpolate values to fit the report cases, and in such instances the cost was prorated on a basis of hearth area of the furnaces. The Nichols data were based on the furnace requirements for handling 30 percent solids in sewage sludge. In our cases, the 30 percent contaminants are dissolved solids and cannot, therefore, be fed directly to the furnace. A flash dryer is provided to do the evaporation required to produce a solid feed to the furnace, but the furnace cost per unit capacity was not reduced corresponding to the reduction in this evaporation load because it was believed that some allowance should be made for the necessity of designing the furnace for continuously firing auxiliary fuel.

It is thought that these two elements will approximately balance one another.

The cost of the flash dryers was derived from Chilton (Reference 10, page 159), the curve values being reduced by 18 percent for drying without disintegration. No further deduction was made for elimination of the furnace normally included in such systems, because this system will have a higher recycle ratio than average. It is thought that these elements will largely offset each other. It is assumed on this basis that flue gases emanating from the flash dryer collector will be sufficiently free from dust that a secondary separation device to prevent atmospheric pollution will not be required. Experimental evidence is required to verify the correctness of this assumption.

Operating costs for incineration were calculated on the same basis as for the evaporation systems in Chapter 3.

Table 2. FLASH DRYER AND FURNACE DESIGN DATA

Case number	1	2	3	4	5	7
Dryer feed 30% solids. 1000 gpd	10.000	254	8.42	100	2.54	1
lb/hr	4.120.000	104.650	3.470	41.200	1.046	412
Flash Dryer (drying to 90% solids)						
Number of units	165	4	1	2	1	1
30% Feed per unit. lb/hr	24.970	26.160	3.470	20.600	1.046	412
Water evaporated per unit. lb/hr	16.650	17.400	2.313	13.730	697	274
Capital cost installed per unit. $	132.500	132.500	44.100	116.000	32.800	32.800
Furnace (drying to 100% solids and oxidizing organics)						
Number of units	165	4	2	2	1	1
Hearth diameter. ft	22.25	22.25	6.0	22.25	6.0	6.0
Hearths per unit	12	12	6	10	6	6
Capital cost installed per unit. $	375.000	450.000	50.000	425.000	50.000	50.000
TOTAL INCINERATOR						
Labor. men/shift	83	2	1	2	0.5	0.25
Labor. cost $/365 days	2.11m	50.800	25.400	50.800	12.700	6.350
Heat. m Btu/hr	907	22	1.19	12.3	0.4	0.170
Heat. cost $/330 days	1.79m	43.500	2.350	24.300	790	336
Power. kw	32.700	792	20.7	329	8.6	3.4
Power. cost $/330 days	1.80m	44.000	1.150	18.300	480	188
Capital. $[a]	96.2m	2.690.000	174.100	1.337.000	97.800	97.800
Fixed charges. $/365 days	10.98m	306.000	19.800	153.000	11.150	11.150
Operating capital cost. $/365 days	0.11m	2.970	330	1.650	169	120
Total annual costs. $/yr	16.79m	447.270	49.030	248.050	25.289	18.145
Unit cost. $/1000 gallons dryer feed	5.08	5.31	17.65	7.52	30.10	55.1
Unit cost. $/1000 gallons evaporated	6.09	6.43	21.20	9.06	36.30	66.2
Unit investment. $/gpd waste	9.62	0.269	0.0174	13.37	0.978	97.8

[a] Includes buildings and miscellaneous not itemized above.

The results of these computations are shown in Table 2. Cases 6, 8, and 9 involve such a small daily amount of incinerator feed (84.2, 25.4, and 0.84 gpd) that it is not practicable to design an incinerator system for them. Indeed the volumes are so small that it would no doubt be cheaper to haul away in these cases the 30 percent liquor or even the 10 percent liquor. However, the interrelations between disposal processes (here hauling vs. incineration) are outside the scope of the present report.

Hauling Waste Solids by Rail

FREIGHT RATE STRUCTURE

Freight rate structures in the United States are extremely diverse and complicated and obviously some simplification is needed for planning purposes. The country is divided into five territories, each having its own rate schedule. Within each territory, rates are not proportional to straight line miles between points, nor even to route miles. Rates also vary tremendously according to the material being hauled. Waste solids hauled in bulk in hopper cars would take the lowest class rate, which is Class 13. Abrupt changes in rates may occur at the boundaries of territories. Also intrastate hauling is not covered by Interstate Commerce Commission regulations, and accordingly, abrupt changes in rate may occur at state lines when the hauling changes from intrastate to interstate.

CLASS RATES VS. COMMODITY RATES

To handle the large variety of materials transported by rail, the class rate system was set up in order that a reasonable number of rate classes could be used to cover all materials. However, if any given material is shipped in large quantities and on a more or less continuous schedule, as for example coal or oil, then "commodity rates" may be set up for that particular commodity between specified points. Commodity rates are much lower than the corresponding class rates. They may be one-third or even less the lowest class rates. Indeed most carload freight moves on commodity rates (e.g. 79% of carload freight revenue in 1953).[22]

If large scale railroad hauling of advanced waste solids

develops in this country, it is likely that freight rates would amount to only one-third or less of those Class 13 rates used in this report. However, there is no way of knowing just what rate would be reached, and furthermore, the commodity rates would be extremely variable from point to point. Also if, as is likely, the waste hauling business is developed in a fragmentary manner, both as to location and time, then commodity rate adjustments will also be fragmentary and not subject to generalization. Accordingly, in the present report the above reservation is made for future application but the computations will be based on Class 13 rates for 1961.

ACTUAL CLASS RATES FOR WASTE SOLIDS

To develop basic freight rates for waste solids, the country was divided into three regions by a series of lines equidistant from sea coasts. This gave a Western district, closest to the Pacific Ocean, bounded by a line traveling northeast from approximately the location of El Paso to a point in southeastern North Dakota; an Eastern district bounded by a northwestward trending line from the middle of Florida to the common point in southeastern North Dakota; and the remainder of the country being in the Midcontinent district. Hypothetical hauls were laid out approximately bisecting each district, and along these lines Class 13 freight rates were obtained for the hauling of waste solids from specified points to Oakland, California; Norfolk, Virginia; and Galveston, Texas, respectively. These data are shown in Table 3. Also shown is the straight line distance from each point to the destination point.

In the last section of the Table are shown rates for short hauls. The ICC prescribed Class 100 rates for 1951 for the short distances used were multiplied by 0.13 to arrive at Class 13 rates. The 1961 rate for 200 miles was computed as 24.6 cents per hundred-weight, which is proportional to the 1961 rate for Austin to Galveston at route miles of 211.8. The remaining distances 0, 100, and 300 miles were adjusted from the 1951 values to 1961 values in proportion to the 1961-to-1951 ratio for 200 miles. The route miles 100, 200, and 300 were changed to hypothetical straightline miles by applying the ratio of straightline miles to route miles for a few of the shorter hauls listed in the Table.

Table 3. FREIGHT RATES FOR WASTE SOLIDS

	Route miles	Straightline miles	¢/cwt	$/ton
Western district [a] To Oakland, Calif., from:				
Bismarck, N. D.	1787	1250	118	23.60
Rapid City, S. D.	1610	1070	118	23.60
Caspar, Wyoming	1488	910	106	21.20
Logan, Utah	847	620	67	13.40
Salt Lake City, Utah	811	590	65	13.60
Reno, Nevada	236	180	34	6.80
Sacramento, Calif.	88	70	24	4.95 [b]
Eastern district [c] To Norfolk, Va., from:				
Minneapolis, Minn.	1309	1040	75	15.00
Milwaukee, Wisc.	999	760	62.5	12.50
Ft. Wayne, Ind.	778	560	54.5	10.90
Columbus, Ohio	632	410	48.5	9.70
Clarksburg, W. Va.	477	280	41.5	8.30
Richmond, Virginia	89	89	20.5	4.10
Midcontinent district [d] To Galveston, Texas, from:				
Duluth, Minn.	1338	1225	76	15.20
Minneapolis, Minn.	1198	1090	71	14.20
Des Moines, Iowa	936	860	61.5	12.30
Kansas City, Mo.	727	675	53.5	10.70
Tulsa, Okla.	518	480	43.5	8.70
Dallas, Texas	290.9	250	30	6.00
Austin, Texas	211.8	175	26	5.20
ICC Class Rates [e]				
	0	0	9.55	1.81
	100	84	18.7	3.74
	200	168	24.6	4.92
	300	253	29.2	5.84

[a] Reference 23.
[b] Includes surcharge at level of wt of shipment = α
[c] Reference 24.
[d] Reference 25.
[e] Reference 26, page 173.

These data are plotted in Figure 3 as a function of straight-line miles. The difference between the Eastern and Midcontinent rates is entirely caused by the more circuitous route necessary along the chosen base line in the Eastern district, which makes

straightline miles a lower percentage of route miles in the Eastern district than in the Midcontinent district. The rates per ton per route mile are practically identical for the two districts (see Figure 5). Along the Western base line route mile rates are about 1 to 1-1/2 cents per hundred-weight per hundred miles greater than for the Eastern and Midcontinent base lines. In addition, the route departs very widely from the straightline route not only because of the mountanious terrain but also because of the scattered population.

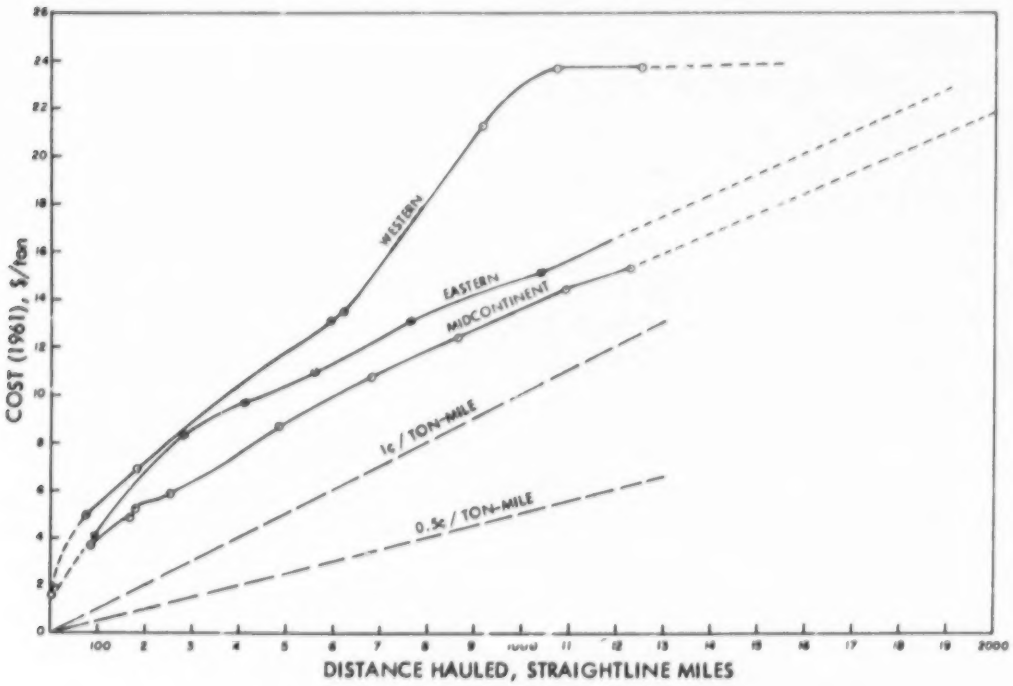

Figure 3. Rail freight rates for waste solids.

THE NATIONAL PATTERN FOR RAIL HAULING

Figure 3 was used to develop the lines of equal hauling costs for waste solids, as shown in Figure 4. It should be understood that these lines are in the nature of "smooth average curves," as a first approximation to the actual conditions. Strictly, they apply only along the sample routes chosen and have been generalized on the basis of straightline miles to cover the entire country. The original distances were those to the nearest (straightline) sea coast, irrespective of whether a railhead existed at that particular coast point. The actual lines of equal cost will be much more contorted than these generalized lines for Class rates because of divergence of route miles from straightline miles, centralization of unloading facilities

at specific coast railheads, and application of group rates (in which all points in a designated geographical area take the same rate; e.g. Bismarck and Rapid City rates in Western district, Table 3). A much greater degree of contortion will arise from application of commodity rates.

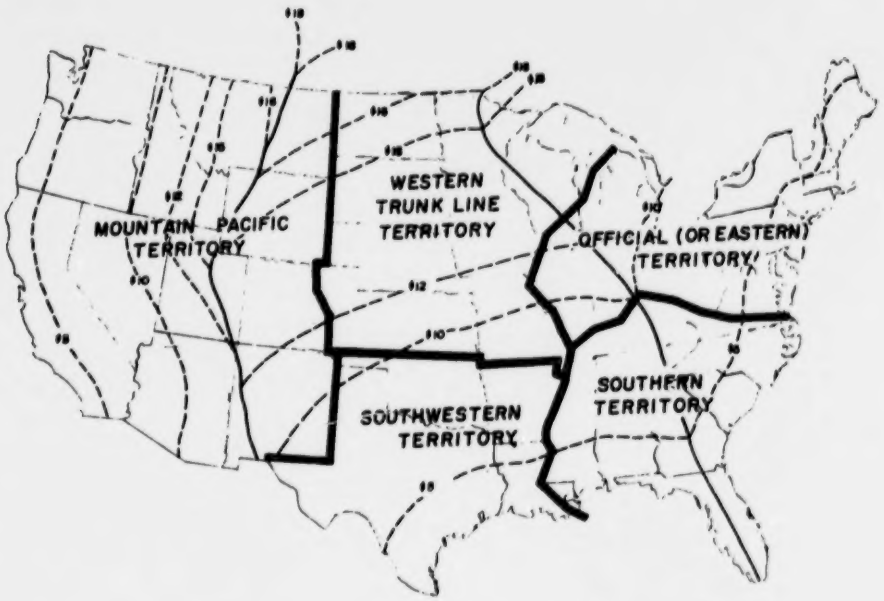

Figure 4. Cost per ton of hauling waste solids by rail (1961 class rates).

Finally, these generalized lines will be further contorted by the operation of competitive rate-setting forces. In that connection it should be cautioned that the line of demarcation between the Pacific Ocean destinations and the Gulf of Mexico destinations is subject to uncertainty because the base chosen for the Midcontinent route bisected the Southwestern Territory and the Western Trunk Line Territory but does not necessarily apply to a line that enters the Mountain Pacific Territory. An abrupt rate change might occur at the boundary of the Mountain Pacific Territory. To illustrate, it is likely that the railroads in the Mountain Pacific Territory would ask for higher rates to the Gulf of Mexico in that portion of the run that occurs in the Mountain Pacific Territory but east of the line of demarcation. This would be done in a competitive effort to move the line of demarcation further eastward in order to increase the business of the roads running to the Pacific. However, to predict these very complicated readjustments is beyond the scope of the present study and Figure 4 will be taken as a first approximation to economic optimum hauling. With that in mind Figure 4 may be taken as showing the

generalized costs of hauling advanced treatment wastes solids to the economically nearest ocean.

LOADING OPERATIONS

In addition to the cost of hauling, there will be incurred costs for loading into the hopper cars. This has been approximated in the following manner with results shown in Table 4.

Table 4. UNIT PLANT COSTS FOR DESIGNED PLANTS

	Dry solids, tons/day	Percent solids	Feed, 1000 gpd	Reported cost million $	Plant cost, $/ gpd
Package plant	1.0	6[a]	8.0	0.1	25.00
Wausau	3.5	5	14.1	0.325	23.00
Wheeling	5.6	6	22.0	0.450	20.45
Chicago	200	3	1,572	11.78	7.48

[a] Value assumed since data were not available.

In Cases 1, 2, and 4, the daily solids to be hauled are so great that direct loading into hopper cars can go on continuously. For Cases 3, 5, 6, and 7, the times between shipments of 40-ton-hopper carloads are from a few days to over a year. Storage would be required for the material while building up to a carload and some rehandling would be required in the loading. The cost of an elevated bin to handle 80 tons of the material has been estimated as about $6000 to $7000 erected, and a loading conveyor brings the total capital to $10,000.

For Cases 8 and 9, the quantities are so small that different handling is indicated. The material would be loaded into 500-pound fiber drums and shipped at LCL rates to the central terminal for ocean dumping. Involved are the cost of the drums and the application of LCL rates, here taken at four times car-

load rates. In an estimation of an individual case, a more detailed time and motion analysis would be made for this operation, but here it is desired to keep the cost parameters within the framework required by the other cases.

Obviously, in an actual situation, for example Case 9 where only 1.63 pounds per day of inorganics are produced, one would not pack this in drums and ship one drum per year to a seashore disposal site. Instead the inorganics would be collected with the floor sweepings or put in the domestic rubbish collection. However, it is desired in the reports in this series to remain within the formal framework of each disposal method without introducing the aspect of optimization among methods or combinations of methods.

As for the loading itself it has been taken that the cost would be about the same as for conveyor loading of bulk materials, namely, about $0.15 per cubic yard (1961).[27] The corresponding line in Table 3 is based on a bulk density of 60 pounds per cubic foot.

Hauling by Other Forms of Transportation

Not only will the pattern and costs of hauling wastes be altered, as suggested in the previous chapter, by interactions between railroads, but are also very likely to be altered in a major way by interactions between railroads and other forms of transportation. For very short hauls up to 50 miles or so, it is likely that truck transportation would be more economical than rail, especially if the daily quantities are small. However, no attempt will be made to estimate the costs of these short hauls in this report because it is very likely that later studies in this series will show that no combination of incineration to dry solids and hauling these short distances can compete in cost with direct pipelining of the liquid or slurry wastes for short distances.

COST OF BARGING

The nation has a well-developed inland waterways system on the Mississippi, Ohio, Missouri, Illinois, and Tennessee rivers, as well as on many of the rivers running directly into the Atlantic or the Gulf. Barge rates are lower than rail rates for two major reasons. First, the actual cost is less since the tractive force is less and the quantities are generally larger, and second, barging does not bear the cost of the waterway as rail hauling

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must bear the cost of the railway. In some cases contract rates for barge hauling are so low that no railroad rates can approach them. For example, in 1942 contract rates on the Great Lakes for grain were about 0.1 cent per ton-mile, for iron ore 0.08¢ and for coal 0.05¢.[28] On the Mississippi river system contract rates in cents per ton-mile were in 1942: for coal, 0.23; petroleum, 0.285; grain, 0.21; and sugar, 0.36. The relation of these barge rates to railroad Class rates is shown in Figure 5.

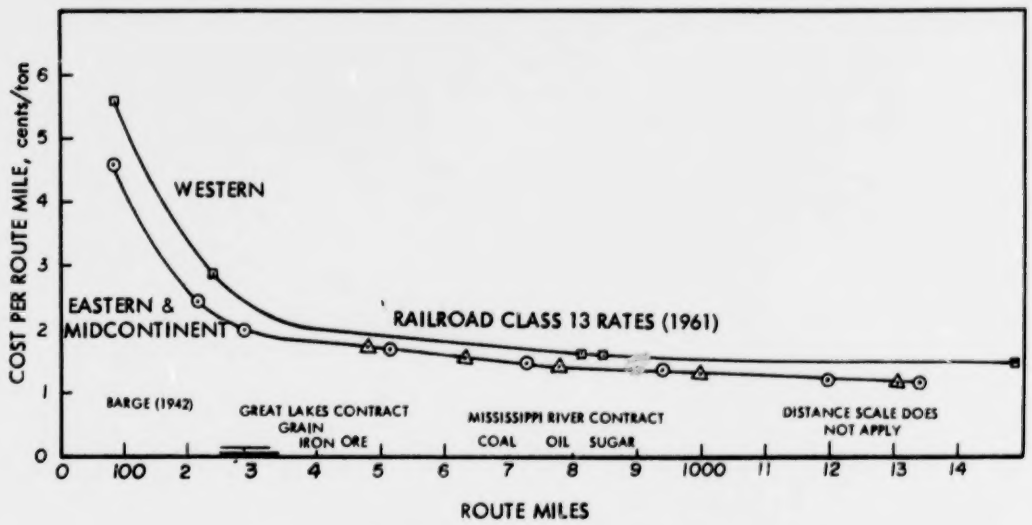

Figure 5. Railroad class rates vs. barge contract rates.

It is of course not fair to compare 1942 barge rates with 1961 railroad rates nor contract barge rates with railroad Class rates. There has undoubtedly been an increase in barge rates since 1942. Common-carrier barge rates are higher than contract rates but always maintain a differential below corresponding railroad rates. On the other hand, it has been pointed out that railroad commodity rates might be as low as one-third of Class rates, so that in the most favorable combination of circumstances it might be that the differential between barging and rail hauling might be decreased considerably more than implied by the above figures. However, on the inland waterways it has been the practice for the common carriers to maintain barge rates at 80 percent of the corresponding rail rates. (Reference 26, page 729).

Furthermore, prominent in any concept of a national waste disposal plan should be that portion of waterway costs not borne directly by the carriers but through taxes upon the nation as a whole. Studies have shown that this "hidden" cost can be equal to the direct carrier rate and in some cases very much more. Reference 29 reports studies of the "Federal

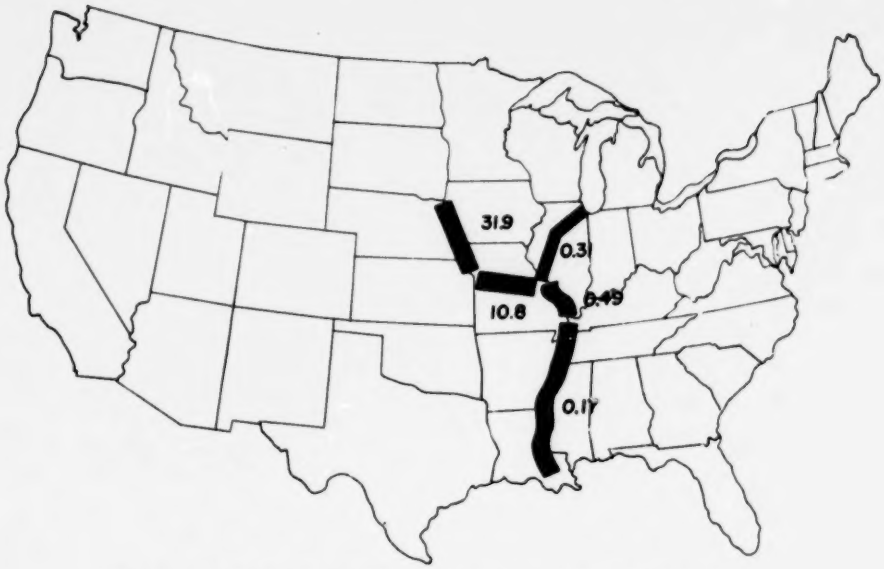

Figure 6. Federal costs in cents per ton-mile on various portions of Mississippi River System.

costs" of maintaining certain waterways in relation to the volume of all traffic carried thereon. Figure 6 shows these costs for portions of the Mississippi system. When these costs are added to the carrier rates described above, they in some cases exceed even railroad Class rates. Of course these "Federal costs" are largely fixed costs and thus added traffic on the waterways would decrease the per ton-mile Federal Cost for transportation of all commodities in the proportion that the total ton-miles carried is increased by the added traffic.

Enough has been indicated by this trend of thought to conclude that economic hauling of advanced treatment wastes can become a very complicated subject worthy of a major transportation investigation of its own not possible within the scope of the present study.

RAIL AND BARGE PATTERNS

On the basis of purely qualitative information, one may assay some generalized shipping patterns on the assumptions that barge shipment would cost significantly less per ton-mile than rail shipment, and that accordingly, in those areas where the differential was greater than the added cost of the transshipment, it would be most economical to utilize a combined rail and barge system. One must expect the customary fringe areas of conflict where railroads will attempt to adjust rates to make all-rail shipment competitive over rail and barge.

Figure 7 has been developed to show in a qualitative way how the national hauling pattern would be altered by introduction of

Figure 7. Shipping pattern in ice-free season.

barge transport. There would be a large area in the center of the country in which the Mississippi River system would supply the barge traffic, with railroads used to haul the waste to barge loading points. The distance of rail haul would be increased as the corresponding all-rail distance to the coast increased. The line of demarcation between the Mississippi River system and the Great Lakes - St. Lawrence Seaway system has been drawn with the fact in mind that Great Lakes rates are considerably lower than Mississippi River rates. No definite cost figures can be placed upon this pattern within the scope of the present study. It can be noted, however, that in the areas indicated as rail and barge on Figure 7 the hauling costs will be less than those shown for all-rail on Figure 4.

One of the disadvantages of barge traffic is that it is seasonal in regions where the waterways freeze. For example, the New York Barge Canal is open only 7 months out of the year and the St. Lawrence Seaway is approximately equally icebound. Figure 8 has been prepared to show roughly how the hauling pattern would change in the icebound season with a year-round ice-free point located at the junction of the Ohio and the Mississippi Rivers.

Dumping at Sea

At coast destinations, dump facilities would be provided whereby hopper cars of waste could be unloaded by dumping directly into seagoing barges. There would be some cost,

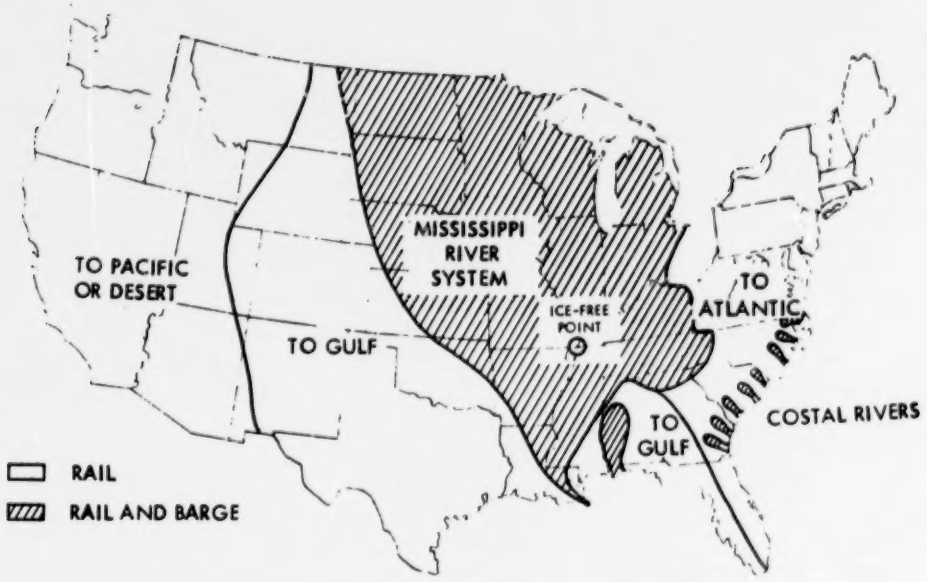

Figure 8. Shipping pattern in ice-bound season.

which is not estimated here, for the maintenance and amortization of these presumably centralized pooled facilities. To eliminate transshipment when inland hauling is by barge, it is suggested that the inland barges be of a seagoing type so that they can be taken directly in tow by seagoing tugs. Unloading the barges at the sea dumping ground would constitute an additional expense not computed here. It is possible that this expense could be reduced to zero by a specially designed barge. Possibly a barge could be constructed with peripheral double walls to supply buoyancy; the enclosed section could be provided with a hopper bottom for dumping directly into the ocean. Possibly this hopper bottom could be designed so that it would be sealed against the entrance of sea water. However, if the seal was not perfect, the operation might still be tolerable since the amount of material lost in transport, i.e. near shore, would be small and consist mostly of the soluble material.

OFFSHORE HAUL DISTANCE

The material being dumped into the ocean is in itself fairly innocuous and for the most part soluble. Indeed, originally, almost all the material entered the process dissolved in water. Some changes may have occurred during incineration to alter the original materials, for example, calcining calcium sulfate

to calcium oxide, and in addition some inherently insoluble material may have resulted from the inorganic components of the incinerated organic matter. In addition to this, there would be whatever ash is residual from direct firing if a direct-fired incineration method is used. Essentially, however, the bulk of the material will be only those salts which would, except for advanced waste treatment, have found their way into the water courses of the nation and thus into the ocean anyway.

The proper distance for hauling these out into the ocean for dumping would have to be determined by oceanographic studies well beyond the scope of the present investigation. For the present computation it was assumed that these solids are barged 30 miles off shore to insure that none of the small amounts of insoluble matter find their way back to the beaches.

COST OF OCEAN BARGING

In the following analysis, the basic data on costs and tug speed have been supplied by Dr. Don Hood, Professor of Chemical Oceanography, Texas A & M College.[30] The assumptions on amorization rates and distances are those of the author.

A seagoing tug averages 5 to 7 knots, thus requiring 11 to 13 hours at $40 per hour to cover the 60 mile round trip, or approximately $500 per round trip. The average barge handles 2000 tons, thus yielding a tug charge of $0.40 per ton.

A barge to handle 2000 tons will cost about $300,000. At an average of 1.5 trips per day, 300 days per year, 10-year amortization, interest at 4 percent, taxes of 1 percent, and insurance 1 percent, this gives $0.047 per ton for the fixed cost on the barge.

The cost of the loading facilities is not estimated here.

The total cost of ocean barging from railhead thus is approximately $0.45 per ton.

The above estimates have been made for rather large capabilities. It is assumed that for the smaller capabilities the solids would be pooled at the barge loading point so as to maintain the low unit cost even at low daily tonnage.

ALTERNATIVES TO BARGING

Because of the nature of the waste, it is conceivable that barging could be dispensed with and the waste could be slurried with sea water and pumped out an outfall line. Sewer

outfall lines from major cities handling primary effluent are in operation. For example, the Hyperion outfalls are 5 and 7 miles long. If sewage sludge and effluent can be handled in this manner, there seems no reason why this much more innocuous waste could not be similarly handled. However, the cost of the ocean barging is so small compared with the probable cost of hauling to the railhead that not much overall saving could be achieved and for this reason the estimate is not made here.

It is also conceivable that the material could be dumped much nearer shore, for example, off the end of a 1- to 2-mile jetty. This would involve detailed oceanographic studies to determine the exact fate of the waste so discharged.

Correlation of Costs

Parameters and costs have been developed for pre-evaporation, incineration, loading, hauling, and ocean dumping. In Table 5 the synthesis of these unit costs into the overall cost for the entire disposal process is shown. Only rail hauling has been considered, but it may be understood that alternative forms of hauling may be cheaper where applicable as discussed previously.

The density of the 10 percent liquor is taken as 1.071 grams per milliliter (8.921 lb / gal), which, corresponding to the other densities, is the density of 10 percent NaCl solution at 68°F.

In Cases 5 and 6, line 10, the cost of saline-water-conversion-type evaporation at approximately 100,000 gpd has been extrapolated from the trend of the data in Figure 1 at $2 per 1000 gallons. The data actually show single-effect vapor compression at $1.10 per 1000 gallons, and further investigation may firm up this latter figure. However, in view of the trend of the other saline water conversion data, $2 was chosen as a conservative cost figure.

In Case 6, line 15, the single-effect evaporator for 200 gpd could not be estimated but the unit cost used was extrapolated from data at higher capabilities. Line 23 shows that the contribution of this portion of the pre-evaporation to the total pre-evaporation cost is small, approximately 6% of the total, so that relatively large errors in extrapolation of line 14 may be tolerated within the limits of error in the estimates.

General inspection of lines 22 to 25 shows that the evaporation to 10 percent constitutes the major portion of the total pre-evaporation cost. It is important to recognize there-

fore that in Cases 5 and 6 this is based on a rather infirm extrapolation of saline water conversion estimates to these low capabilities.

In Cases 6, 8, and 9, line 26 shows an extrapolation of incineration costs of more than $200 per 1000 gallons of feed to the incinerator. In these cases the incinerator capability is too small to arrive at a physical conception comparable with the larger equipment on which an estimate could be based. The extrapolation should be taken to mean that incineration in some type of apparatus could be accomplished to conform with the formal requirements of the scope of work at a cost in excess of $200 per 1000 gallons. Lines 33 to 37 show that for cases 6 and 9 the contribution of the incineration to the total cost is minor, being 9 percent in Case 6 and about 0.7 percent in Case 9. This means that even relatively large errors in the unit cost of the incineration operation would have little influence on the total cost of disposal as estimated. In Case 8 the incineration cost amounts to 17.5 percent of the subtotal disposal cost so that errors in the unit incineration cost would be of larger import in the total cost.

Line 32 shows the cost for rail hauling to the ocean per thousand gallons of waste from those points on the map, Figure 4, from which the unit cost of hauling is $1 per ton. The cost of hauling per thousand gallons of waste is of course proportional to the cost of hauling per ton, as shown on the map; overall costs including the hauling from various $/ton contours of the map are shown in lines 38 through 42.

In Cases 3, 6, and 9, the distance from the ocean is a negligible consideration in the cost of disposal because the quantities of inorganics handled are so small compared to the initial volume of waste. The same is true of Cases 5 and 8, but not for the case corresponding to these, Case 2, because the subtotal cost for Case 2 is already so mall that the hauling cost (which is the same for Case 2 as for Cases 5 and 8) becomes appreciable compared to it. One may therefore say as a first approximation, surprisingly, that the distance from the ocean is of no importance in the unit cost of disposal except when the waste is concentrated. The lower the daily volume, the higher the breakpoint concentration at which unit hauling costs become appreciable.

INCINERATION										
Cost, $/1000 gal of feed to incinerator	26	5.08	5.31	17.65	7.52	30.10	(> 200)	55.1	(>200)	(>200)
Cost, $/1000 gal of oringinal waste	27	5.08	0.135	0.015	7.52	0.765	(>0.21)	55.1	(>6.35)	(>0.21)
Solid product (inorganics) to haul, ons/day tons/day	28	9.660	250	8.32	96.6	2.50	0.0832	0.966	0.0250	0.000832
OCEAN DUMPING										
Tons inorganics per 1000 gal of waste	29	0.966	0.025	0.000832	0.966	0.025	0.000832	0.966	0.025	0.000832
Cost, $/1000 gal of waste of waste	30	0.435	0.011	0.000374	0.435	0.011	0.000374	0.435	0.011	0.000374
RAIL HAULING										
Loading and storage, $/1000 gal	31	0.181	0.0047	0.002	0.181	0.032	0.027	2.85	0.255	0.00834
Cost, hauling to ocean $/1000 gal of waste per $1.00/ton, from map	32	0.966	0.025	0.000832	0.966	0.025	0.000832	0.966	0.10	0.0033
SUMMARY										
Costs per 1000 gal of Waste										
Evaporation to 30%	33	0	0.660	0.541	0	2.57	2.12	0	30.5	31.3
Incineration	34	5.08	0.135	0.015	7.52	0.765	> 0.21	55.1	> 6.4	> 0.21
Ocean dumping	35	0.435	0.011	0.000374	0.435	0.011	0.000374	0.435	0.011	0.000374
Loading and storage	36	0.181	0.0047	0.002	0.181	0.032	0.027	2.85	0.255	0.008
Subtotal	37	5.70	0.811	0.558	8.14	3.39	> 2.35	58.3	> 37.1	31.5
TOTAL:										
with rail haul of $ 0/ton	38	5.70	0.8	0.558	8.14	3.39	> 2.35	58.3	> 37.1	> 31.5
with rail haul of $ 1.00/ton	39	6.67	0.836	0.559	9.11	3.41	> 2.35	59.3	> 37.2	> 31.5
with rail haul of $ 5.00/ton	40	9.53	0.936	0.561	12.97	3.53	> 2.35	63.1	> 37.6	> 31.5
with rail haul of $ 10.00/ton	41	15.36	1.061	0.565	17.80	3.66	> 2.35	68.0	> 38.1	> 31.5
with rail haul of $ 17.00/ton (approximates U.S. maximum)	42	22.10	1.236	0.571	24.54	3.84	> 2.36	74.7	> 38.8	> 31.5
$/day disposal cost at $10 haul (approximates U.S. average)	43	153,600	10,610	5,650	1,780	366	> 235	68.0	> 38.1	> 31.5
at zero haul	44	57,000	8,110	5,580	814	339	> 235	58.3	> 37.1	> 31.5
UNIT INVESTMENT COST										
Approximate, $/gpd	45	9.62	1.92	1.53	13.37	4.72	3.18	107.8	18.8	18.8

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Table 5. THE DISPOSAL COMPONENTS AND COSTS

	Line no	Case Number 1	2	3	4	5	6	7	8	9
Feed rate, 1000 gpd	2	10,000	10,000	10,000	100	100	100	1	1	1
Organics, ppm	3	100,000	3,000	100	100,000	3,000	100	100,000	3,000	100
Inorganics, ppm	4	200,000	6,000	200	200,000	6,000	200	200,000	6,000	200
Total contaminants, ppm	5	300,000	9,000	300	300,000	9,000	300	300,000	9,000	300
Density, g/ml	6	1.187	1.005	1.000	1.187	1.005	1.000	1.187	1.005	1.000
Tons/day contaminants	7	14,850	377	12.5	148.5	3.77	0.125	1.485	0.0377	0.00125
PRE-EVAPORATION										
Evaporation to 10%										
Water evaporated, 1000 gpd	8	none	9,[illegible]46	9,970	none	91.46	99.7	none	0.915	0.997
Process used	9		SWC	SWC		SWC	SWC			
Unit cost evaporated water, $/1000 gal	10		0.54	0.53		(2.00)	(2.00)			
Rate, 1000 gpd	11		8.45	28.1		8.45	0.281		0.0845	0.00281
Evaporation from 10% to 30%										
Water evaporated, 1000 gpd	12	none	60[illegible]	20.2	none	6.04	0.2	none	0.0604	0.002
Process used	13		tripleeffect	triple effect		triple effect	single effect			
Cooling water, 1000 gal/1000 gal of waste	14		0.494	0.0166		0.49	(0.05)			
Unit cost, evaporated water, $/1000 gal	15		2.77	6.64		12.25	(66.0)			
Evaporation direct to 30%										
Water evaporated, 1000 gpd	16							none	0.975	0.999
Process used	17								single effect	single effect
Cooling water, 1000 gal/1000 gal of waste	18								24.0	24.6
Unit cost, evaporated water, $/1000 gal	19								31.3	31.3
Rate, 1000 gpd 1000 gpd	20	10,000	254	8.42	100	2.54	0.0842	1	0.0254	0.000842
Rate, lb/day ·ate, lb/day	21	98.88×10^6	2,511,600	83,300	988,800	25,116	833	9,888	251	8.33
Cost per unit original waste feed										
Evaporation to 10%, $/1000 gal	22		0.494	0.528		1.83	1.99			
Evaporation from 10 to 30%, $/1000 gal	23		0.166	0.013		0.735	0.132			
Evaporation to 30%, $/1000 gal	24								30.5	31.3
Total evaporating cost, $/1000 gal	25		0.660	0.541		2.57	2.12		30.5	31.3

Line 45 shows an approximation of unit investment cost per gpd. Cases 1, 4, and 7 are of the same accuracy as the estimates in the report. But Cases 2, 3, 5, and 6 involve as the major component an arbitrary investment for the first stage of the pre-evaporation, the saline water conversion type, of $1.50 per gpd at 10 mgd and $3.00 per gpd at 100,000 gpd, and reflect largely the values assumed for this component. Case 6 involves a crude estimate of the unit cost of evaporation from 10 to 30 percent, which amounts to only about 2 percent of the total investment. Cases 8 and 9 involve a crude estimate of incinerator cost, which amounts to only about 5.5 percent of the total investment.

COMPARISON WITH WET OXIDATION AND INJECTION

A correlation according to capability is shown in Figure 9 where a comparison is made with unit costs of wet oxidation and of injection (A report on injection will be published later). Incineration unit costs are shown both for the U.S. average haul of about $10 per ton and for zero haul. At 300 ppm the zero haul data are indistinguishable on the Figure from the average haul data. Comparison of the distance between the zero haul and average haul for 9,000 ppm with that for 300,000 ppm reiterates what has been stated, namely, hauling costs decrease in importance as concentration decreases.

Also shown on the Figure are costs for average conditions for wet oxidation and injection (References 1 and 2). Again it

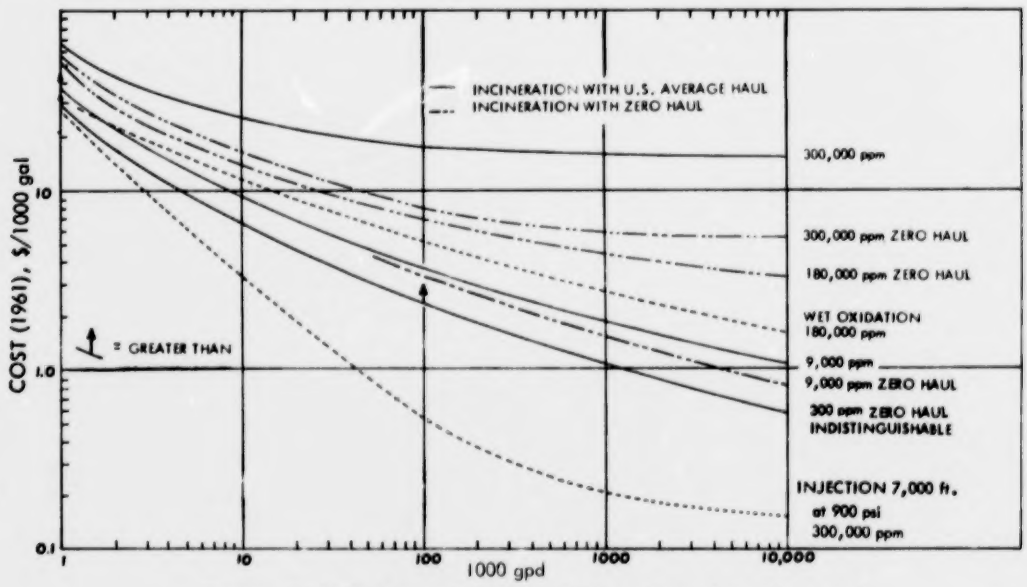

Figure 9. Unit disposal costs by incineration, injection, and wet oxidation.

is seen that the unit costs for injection are considerably lower than those for the other methods.

It is interesting to compare, as in Figure 9, incineration with wet oxidation. Both processes accomplish the oxidation of the organic matter, wet oxidation process to the extent of about 80 percent and incineration presumably higher. Wet oxidation leaves the residue plus the unreacted inorganic contaminants in an aqueous form whereas incineration reduces them to ash. In the wet oxidation calculations, no provision was made for ultimate disposal of the resulting liquor. Therefore, among the available possibilities, it seems most appropriate to compare it with incineration at zero haul, that is with no cost for hauling the residue in either case. The wet oxidation under the conditions chosen had an upper limit of 180,000 ppm. The incineration costs compared in Figure 9 have been interpolated to that concentration for comparison; the cost of incineration at zero haul is greater than the cost of the corresponding wet oxidation at all capabilities. The possibility exists, and should be explored, that pre-evaporation for the operation might be carried out as a wet oxidation.

ECONOMIC CHOICE

In Figure 10 data are provided for economic choice between incineration and injection. Figure 10 corresponds with Figure 15 in Reference 2.

The black lines in the Figure refer to ultimate disposal by incineration at the U.S. average haul of $10 per ton. The red lines superimposed refer to injection at an average depth of 7000 feet and at 900-psi standard wellhead pressure, which is the 90 percent percentile pressure of that study. The approximately horizontal red lines show that daily disposal costs for injection are practically independent of the tons per day of solids disposed and are almost totally a function of the volume disposed. However, the corresponding black lines for incineration at constant daily volume show increasing daily disposal costs as the concentration increases. That some, but not most of this increase is caused by the average haul chosen, is shown by the zero-haul lines, which show the corresponding daily disposal cost with zero haul.

The slopes of the lines of equal concentration for incineration are similar to the slopes of the corresponding red lines for injection.

Comparison of the constant capability lines shows that for the cost of incinerating 1,000 gpd, one could inject 10,000 to 100,000 gpd. For the cost of incinerating 100,000 gpd, one could inject 1 to 10 mgd.

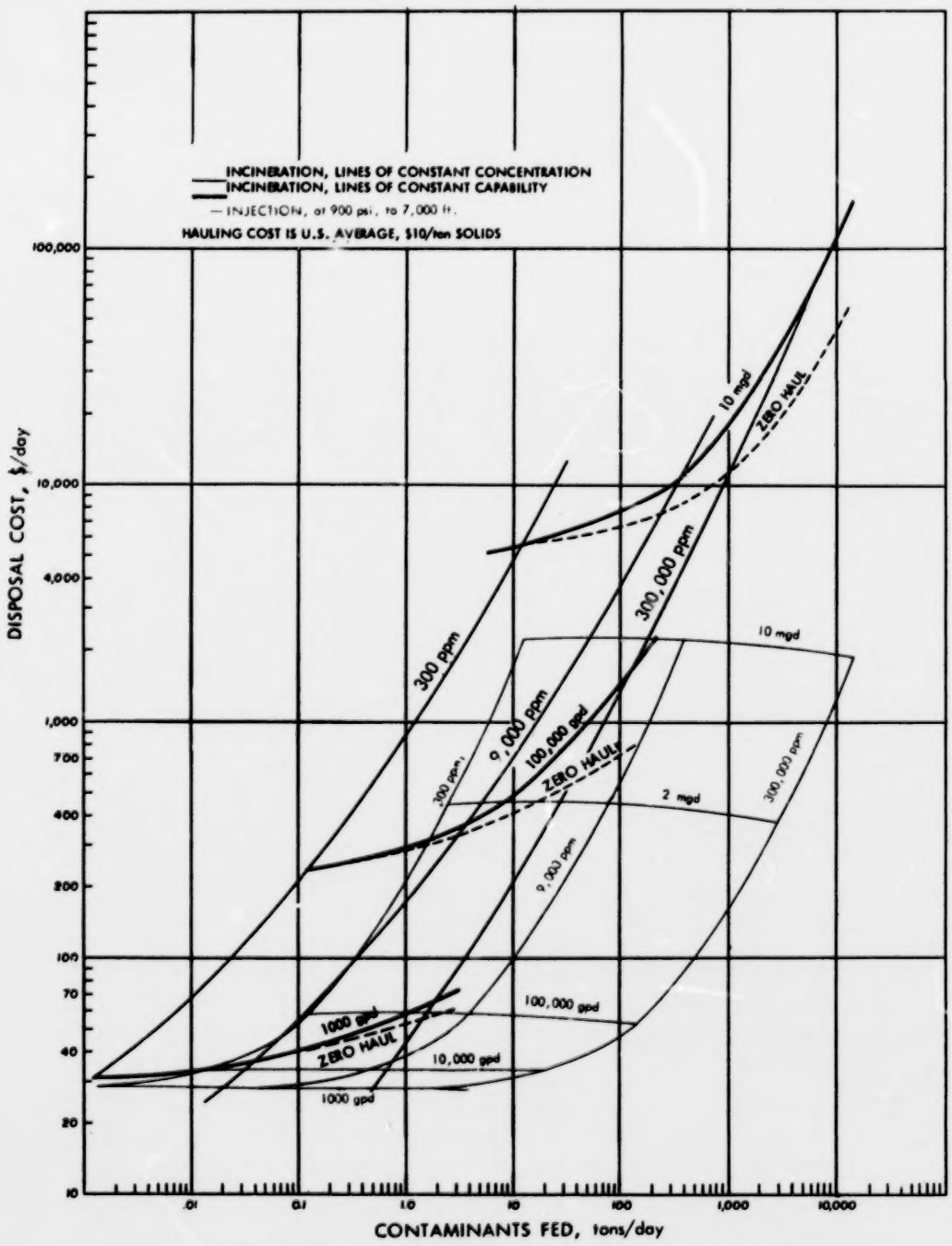

Figure 10. Economic comparison of incineration and injection.

LOGISTICS

Table 6 and Figure 11 show disposal costs in $/mg of water produced in advanced treatment as a function of the percent

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Table 6. DISPOSAL COST AS A FUNCTION OF PERCENT RECOVERY

Recovery of product water in advanced treatment, %	Contaminants in advanced treatment waste, ppm	Disposal cost, $			
		Per mg of water produced in advanced treatment			Per mg of total water produced
		by wet oxidation	by injection	by incineration	by incineration
99.90	300,000		3.20	24.0	24.0
99.83	180,000	15.5	3.80		
96.67	9,000	91.1	10.5	82.5	80.1
80.00	1,500	538	50.0	240	192
50.00	600		196	670	335

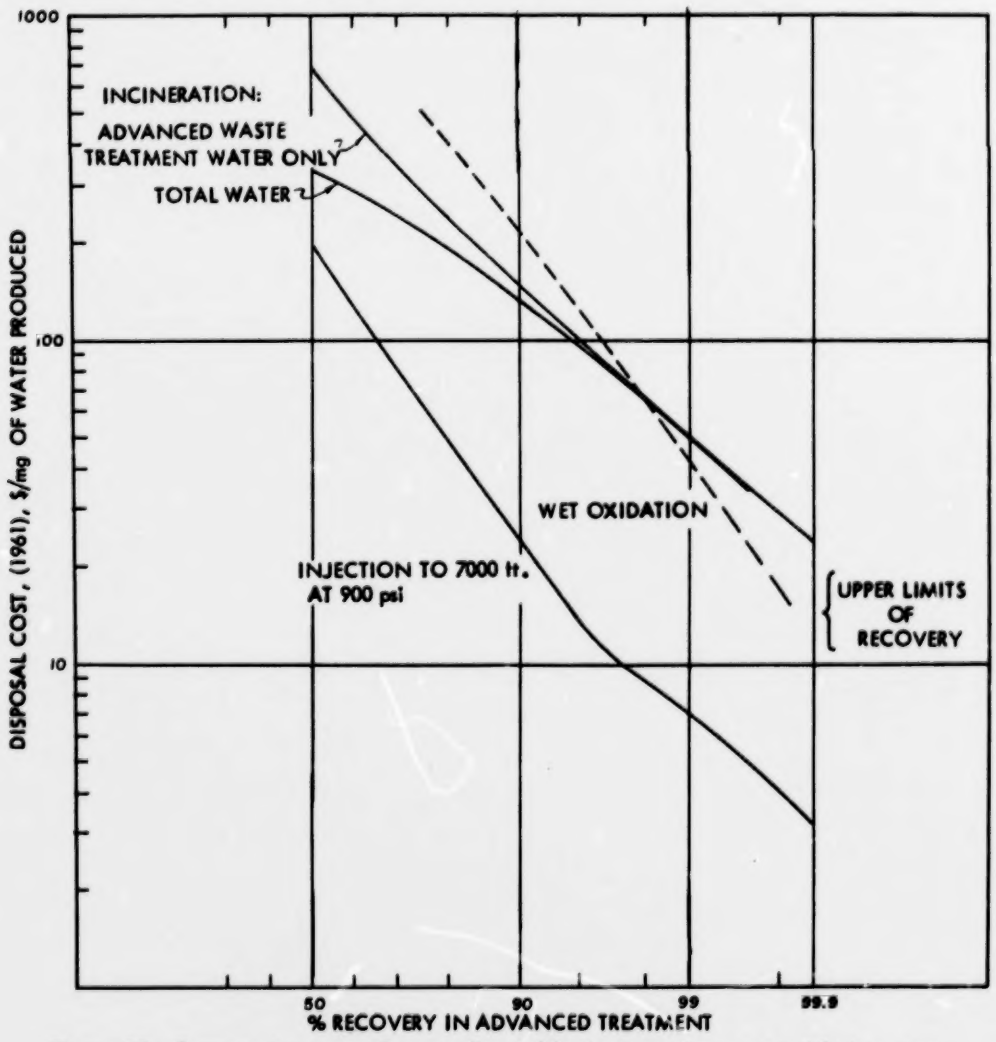

Figure 11. Comparative costs of waste disposal by incineration, wet oxidation, and injection (10 mgd of secondary effluent containing 300 ppm contaminants).

recovery in advanced treatment. Figure 11 corresponds to Figure 16 in Reference 2.

It is shown that the cost of incineration at average hauls is roughly equal to the cost of wet oxidation for the illustrative

case chosen, namely, 10 mgd of a secondary effluent containing 300 ppm contaminants (200 ppm inorganic and 100 ppm organic). Injection under average conditions is much less expensive per million gallons of water produced than either of the other methods.

The two lines for incineration refer to the interesting circumstance that the pre-evaporation step will produce distilled water practically free from the non-volatile constituents and possibly containing some of the volatile contaminants in the original waste. Information is not yet available on the non-volatile content of water distilled from secondary effluent, but if it be assumed that this water is potable, then the total production of water from the advanced waste treatment plus the disposal operation will be increased to equal a total water recovery of 99.90 percent. The costs per million gallons of this total produced water are shown as the lower line of the incineration pair. This increase in production and concomitant decrease in cost is of little magnitude at high recoveries, e.g. 90 percent and above. However, at 50 percent recovery in advanced waste treatment the water produced in pre-evaporation is itself almost the remaining 50 percent and thus the production of water overall is doubled and the unit cost is approximately halved.

Areas for Further Investigation

Since the estimates in this report are predesign estimates, they obviously will be improved in accuracy by replacement of the necessary assumptions with actual facts determined experimentally or otherwise and by design optimization.

SALINE WATER CONVERSION COSTS

In the pre-evaporation estimates the outstanding assumption is the figure of $2 per 1000 gallons for the cost of saline-water-conversion-type evaporation at capabilities of about 100,000 gpd. The triple-effect evaporation estimated in this report would have a cost of about $3.75 per 1000 gallons at 100,000 gpd, whereas vapor compression plant estimated for saline water conversion showed $1.10 per 1000 gallons. The 10-effect LTV plus vapor compression would extrapolate, if in alloy, to about the same as the triple-effect at this capability. However, the latter is a 10-fold extrapolation of the data. The $2 per 1000 gallons somewhere less than midway between these two figures was selected somewhat arbitrarily. However, it does have an

important part in the water costs in two of the selected cases and for that reason should be placed on a firmer basis.

FOULING AND SCALING

The other serious deficiency of the pre-evaporation design is the lack of knowledge concerning scaling and fouling. In the designed evaporators the maximum liquor temperature has been taken at 220°F, consistent with non-scaling conditions in sea water evaporation up to about 10 percent concentration. However, in the present instance pre-evaporation to 30 percent is planned. The compositions of advanced treatment waste will vary over a wide range and will be different from the compositions found in sea water so far as the inorganics contained are concerned. Some of these compositions will undoubtedly be more subject to scaling than is sea water. Furthermore, in the advanced waste contaminants there is a large organic component whose composition and behavior cannot be predicted. Indeed, the knowledge of the properties of the advanced treatment waste is vague. The organic content at 10 percent organics and 20 percent inorganics might possibly remain in solution, but more likely will be found as a second phase and might be resinous, tarry, or recalcitrant in other ways. Thirty-five days of shutdown per year have been allowed partly for scale removal operations, but it cannot be known whether this is adequate until the nature of the waste can be specified more fully. Obviously, this calls for study of advanced treatment wastes as to their actual compositions and properties in evaporation. Many other factors depend on these properties as well, for example, construction materials, boiling point elevations, viscosities, and other parameters of importance for evaporation.

Some of the same uncertainties regarding the composition of the material being handled apply to the flash dryer and furnace operation in incineration. Whether the apparatus for mixing the 30 percent feed with recycled flash-dried solids will operate without plugging and binding is not completely certain, and the same applies to the flash-drying portion of the apparatus. The furnace itself with its mechanical rabble arms presumably can comminute and carry forward over the hearths, almost any type of solid material with the exception of one that forms a heavy tar prior to oxidation, but this also requires experimental verification.

The Atomized Suspension Technique (AST) could handle these fouling problems without trouble, but as indicated in the text,

a major investigation is necessary to determine whether the AST apparatus can be made to handle saline materials without rapid scaling.

COOLING WATER

The amount of cooling water required compared to the amount of waste feed to the disposal process is large in those cases where pre-evaporation is necessary. Cooling water use is also large in wet oxidation.[1] For these cases then cooling water use becomes of logistical importance in the sense that the present designs and estimates apply to situations in which there is abundant cooling water, as distinct from the supply of advanced treatment feed or waste. This point should be kept in mind when incineration and wet oxidation in arid areas are considered. However, it is not likely that changing to air cooling would greatly increase the overall costs. The cooling water use indicated does not include the cooling water used in the saline-water-conversion-type stage of pre-evaporation.

BALANCING INPUT AND OUTPUT

In an integrated process consisting of two stages of pre-evaporation followed by incineration, it is very likely that utilization of heat would be most efficient if the output from a preceding stage was not cooled to ambient temperature but was used as hot feed to the next stage. However, the saline water conversion costs utilized for the first pre-evaporation stage were estimated on the basis of blowdown at ambient temperatures so that all possible heat was used in that stage. The evaporators for the second stage of the pre-evaporation, as designed in the present study, do not recover heat from the bottoms. A slight decrease in total cost would be achieved if the second stage of pre-evaporation was integrated with the first saline-water-conversion-type stage, with respect to output-input temperatures. A more detailed design and cost study would take this into account.

CONDENSATE QUALITY

It has been suggested that the condensate from pre-evaporation might be potable and suitable for blending with the water product from advanced waste treatment itself, and that under certain combinations of the parameters, would constitute a significant part of the total water produced. Before this can be firmly

established it must be known that the volatile contaminants will not appear in the condensate to an extent that will make it non-potable when blended.

AIR POLLUTION

With the feed material sprayed on recycled solids, it is planned that the particle size of the dried material will be rather large so that extreme measures will not be necessary in cleaning up the exit gas. This is not certain, however. Also there may be some noxious gases resulting from calcination of some of the inorganic solids, e.g., SO_3 from $CaSO_4$. Scrubbers would be required to remove these.

SOLIDS HANDLING

The temperature in the furnace will calcine any calcium or magnesium carbonates that may be present in the waste, and since calcium bicarbonate is a very likely constituent of secondary effluent, it may be expected that active lime will be present in the residue. This may present some problems in handling the solids for hauling and particularly in long-term storage when many days pass between carloadings.

OTHER INCINERATION APPARATUS

The box furnace and the AST reactor have been mentioned as possibilities for incineration. Also, there have been described the nuclear waste rotary ball calciner and the fluidized-bed calciner both of which could presumably be operated at temperatures high enough to insure incineration. It is unlikely that any developmental work will take place on the box furnace unless it is specifically undertaken for the Advanced Waste Treatment project. However, other interests are furthering the development of the AST, as well as the rotary ball calciner and the fluidized-bed calciner. Accordingly, surveillance should be maintained over developments on these three processes in case at some time in the future they might become sufficiently developed to warrant investigation for advanced waste treatment incineration.

UNIT INVESTMENT

The unit investment cost for the saline-water-conversion-type pre-evaporation stage was not readily available, and therefore an overall unit investment cost cannot be developed.

Unit investment costs are of much less importance in a preliminary cost estimation than product costs, but in the present study this does remain as a deficiency that should be made up in any more detailed study.

TRANSPORTATION ECONOMICS

In developing the costs of hauling waste solids by rail, a high degree of smoothing was used in choosing only a single bisector for actual rate data and in changing route miles to straightline miles. The next stage in a study of this subject would be to determine actual rates at a large number of specific originating points throughout the country and presumably to a relatively small number of seashore collecting points. Map contours thus developed would result in some changes in the boundaries between Pacific, Gulf, and Atlantic areas. Much more important however would be the application of commodity rates rather than class rates. Also involved is the interaction between rail rates and barge and trucking rates. Such a study would be of a magnitude that would disassociate it entirely form the engineering study, and presumably should be carried out by a specialist in transportation economics.

MARITIME AND OCEANOGRAPHIC STUDIES

In ocean dumping the cost of the seashore facilities should be included in any future estimates. An oceanographic study is called for to determine the proper offshore distance for dumping, and considerable maritime engineering type of study could be involved in developing the proper type of ocean barge for hauling the solids. Other outfall methods such as outfall lines are possiblities and could be made the subject of a specialized study.

COMBINATION PROCESSES

The above comments on areas for further investigation are, as with the other reports in this series, restricted to the subject of the report itself. Obviously, there are numerous areas for further investigation among combinations of disposal processes. In this investigation, a combination process (evaporation plus incineration) was studied because it was judged and then proven that evaporation in an evaporator would be less expensive than evaporation in an incinerator. There

are also other forms of combinations involving, for example, whether to pipeline or tank car pre-evaporated waste in preference to incinerating it. All of these combination investigations must remain for later activation after the pure forms of disposal have been estimated; an approach that introduced combination techniques at too early a stage would lead only to chaos.

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ERRATA

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Table 4 on page 23 should appear on page 53, and Table 4 on page 53 should appear on page 23.